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Effect of Emission Control Requirements on Fluidized-Bed Boilers for Industrial Applications: Preliminary Technical/Economic Assessment

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EFFECT OF EMISSION CONTROL REQUIREMENTS ON FLUIDIZED-BED BOILERS FOR INDUSTRIAL APPLICATIONS:

Preliminary Technical/Economic Assessment

by

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16. ABSTRACT

The report gives results of an assessment of coal-fired industrial fluidized-bed boilers (FBBs) to estimate the impact, on boiler cost and performance, of alternative control levels for SOx, NOx, and particulate emissions. Base industrial FBB conceptual designs are presented, and the cost, boiler plant energy efficiency, and environmental performance are estimated as a function of boiler size, coal characteristics, sorbent properties, and boiler design and operating conditions. One major conclusion is that, even at the most stringent control levels examined in this study, FBBs are competitive with equivalent conventional coal-fired industrial boilers with stack gas cleaning (assuming that the fluidized bed is designed to provide adequate gas residence time in the bed). A second conclusion is that several design and operating options, including increased gas residence time, are available that can promote the potential advantages of industrial FBBs. This study was conducted to provide background information for an EPA effort to develop New Source Performance Standards for air emissions from industrial boilers.

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Assessments	Pollution Control	13B	14B
Fluidized Bed Processing	Sulfur Oxides	Stationary Sources		07B
	Nitrogen Oxides	Industrial Boilers	13H	
Coal	Dust	Particulate	21D	11G
Combustion			21B	
Boilers			13A	

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PREFACE

The Westinghouse R&D Center is carrying out a program to provide experimental and engineering support for the development of fluidized-bed combustion systems under contract to the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency (EPA), at Research Triangle Park, NC. The contract scope includes atmospheric and pressurized fluidized-bed combustion processes as they may be applied for steam generation, electric power generation, or process heat. Specific tasks include work on calcium-based sulfur removal systems (e.g., sorption kinetics, regeneration, attrition, modeling), alternative sulfur sorbents, nitrogen oxide emissions, particulate emissions and control, trace element emissions and control, spent sorbent and ash disposal, and systems evaluation (e.g., the impact of new source performance standards on fluidized-bed combustion system design and cost).

In a previous report (EPA-600/7-78-163, August 1978), Westinghouse assessed the impact of alternative SO_2 control levels on the technical and economic potential of fluidized-bed boilers in electric utility applications. This earlier study was conducted in support of the then continuing effort within EPA to develop New Source Performance Standards for air emissions from utility boilers. The current study, covered in this report, extends that earlier effort to include the effect of SO_2, NO_x, and particulate control levels for fluidized-bed boilers in industrial applications. This current effort is in support of the continuing EPA effort to develop air emission standards for industrial boilers. Some of the information developed in this report is also utilized in the "Technology Assessment Report for Industrial Boiler Applications: Fluidized Bed Combustion" (EPA-600/7-79-178e, November 1979), prepared by GCA/Technology Division.

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This document contains the results of a technical effort, defined and completed under a technical directive, from January to September 1979. The work reported represents a preliminary application of prior work completed by Westinghouse under contract to EPA.

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ABSTRACT

An assessment of the impact of Alternative Control Levels (ACL) for sulfur oxides (SO_x), nitrogen oxides (NO_x), and particulate emissions from coal-fired industrial fluidized-bed combustor steam generators on cost and performance has been performed. Base industrial fluidized-bed boiler conceptual designs are presented, and the cost, boiler plant energy efficiency, and environmental performance are estimated. Boiler capacity, coal characteristics, sorbent properties, and the fluidized-bed boiler design and operating conditions are parameters in the evaluation. This study considers conventional, dense-bed FBC concepts. High-velocity, circulating-bed concepts are not included in this evaluation. The major conclusions developed are that:

- The proposed control levels examined do not limit the competitive position of the industrial fluidized-bed boiler with respect to the conventional coal-fired industrial boiler with stack-gas cleaning. This conclusion is based on the assumption that the fluidized-bed boiler is designed to provide adequate gas residence time in the bed (≥ 0.67 s).
- Several design and operating options, including increased gas residence time, are available that permit the promotion of the potential advantages of industrial fluidized-bed boilers, even for the most stringent control levels considered in this study.

This study was conducted in order to provide background information for the EPA effort to develop New Source Performance Standards (NSPS) for air emissions from industrial boilers.

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NOMENCLATURE

ACL - alternative control level(s)
AFBC - atmospheric pressure fluidized-bed combustion
ANL - Argonne National Laboratory
Ca/S - calcium-to-sulfur ratio
CTIU - component test integration unit
DWS - drinking water standards
EPA - Environmental Protection Agency
EPRI - Electric Power Research Institute
FBC - fluidized-bed combustion
FGD - flue gas desulfurization
FW - Foster-Wheeler Engineering Corporation
METC - Morgantown Energy and Technology Center
NCB - National Coal Board
NSPS - new source performance standards
PER - Pope, Evans and Robbins, Inc.
PFBC - pressurized fluidized-bed combustion
RCRA - Resource Conservation and Recovery Act
TDS - total dissolved solids
TG - thermogravimetric
TOC - total organic carbon

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1. INTRODUCTION

The fluidized-bed combustion (FBC) of coal represents a developing technology of great promise in industrial steam generation applications. The Department of Energy (DOE) is currently funding several programs to demonstrate the technology in a variety of applications. Several vendors are now also marketing industrial FBC steam generators, although some aspects of the technology are not yet commercially perfected.

The major advantages of the FBC steam generator over the conventional industrial steam generator generally are claimed to be its smaller combustor size, its greater fuel flexibility (liquid, gas, solids, waste materials), and its greater adaptability to environmental emissions restrictions. The potential for shop fabrication of the industrial FBC steam generator is a benefit resulting from its compact dimensions. The interrelation of these potential advantages is not currently well understood.

EPA is currently involved in the development of New Source Performance Standards (NSPS) for industrial steam generators. The objective of this study is to assess the impact of alternative control levels (ACL) that might be considered in the development of NSPS on industrial FBC steam generator performance and cost. A product of the study is a perspective for developers and vendors on the design and operating parameters dominant in setting the performance and cost, the advantages of the industrial FBC steam generator, and the ranges of design and operating parameters that should be incorporated into development programs to ensure that industrial FBC steam generators can economically satisfy future NSPS.

In a previous report (EPA-600/7-78-163, August 1978), Westinghouse assessed the impact of alternative SO_2 control levels on

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the technical and economic potential of fluidized-bed boilers in electric utility applications. This earlier study was conducted in support of the then continuing effort within EPA to develop New Source Performance Standards for air emissions from utility boilers. The current study, covered in this report, extends that earlier effort to include the effect of SO_2, NO_x, and particulate control levels for fluidized-bed boilers in industrial applications. This study considers conventional, dense-bed FBC concepts. High-velocity, circulating-bed concepts are not included in this evaluation. This current effort is in support of the continuing EPA effort to develop air emission standards for industrial boilers. Some of the information developed in this report is also utilized in the "Technology Assessment Report for Industrial Boiler Applications: Fluidized Bed Combustion" (EPA-600/7-79-178e, November 1979), prepared by GCA/Technology Division.

This evaluation incorporates the results of previous engineering and modeling studies in developing preliminary performance and cost projections for industrial FBC steam generators. This work was performed from January to September 1979 in response to a technical directive from the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency (EPA).

2. SUMMARY AND CONCLUSIONS

The objective of this study is to determine the extent to which the performance and cost of industrial FBC are influenced by various ACL. The report attempts to develop valid trends in behavior for this purpose, although the uncertainties in absolute performance and cost factors are probably too great to permit absolute comparisons between industrial FBC and conventional boilers to be meaningful. This study considers conventional, dense-bed FBC concepts. High-velocity, circulating-bed concepts are not included in this evaluation.

The results of the study of the impact of ACL on the performance and cost of industrial fluidized-bed combustor steam generators (referred to henceforth as industrial FBC boilers) is reported in twelve sections. Sections 3 through 5 provide overall perspective on the study, summarizing the scope of the study and the base FBC boiler design specifications and study parameters and discussing industrial AFBC concepts. Section 6 looks at the impact of the NSPS for sulfur oxide (SO_x) emissions on the consumption rate of sorbent. Section 7 develops perspective on the critical area of sorbent particle attrition in fluidized-bed combustion and relates attrition to its effect on industrial FBC boiler performance. Section 8 discusses particulate control in the industrial FBC boiler, with emphasis on operating parameters and design options. Solid waste emission from the industrial FBC boiler is evaluated in Section 9, with specific projections of the rate of total waste solids presented as a function of the ACL. The environmental nature of the solids is also discussed. Section 10 discusses the ability of the industrial FBC boiler to satisfy various levels of control for nitrogen oxides (NO_x). The last two major sections (11 and 12) present projections of total plant energy efficiency and cost, respectively. Again, the impact of ACL is indicated, as is the influence of key design options and operating parameters.

In each section the performance of a base industrial FBC boiler design, with shop-fabricated modules and parametric modifications to that design, are discussed. Only dense-bed FBC concepts, representing the most nearly commercial designs, are evaluated here. The dilute-phase recirculating concepts are not dealt with. When applicable, comparisons are made in each section between industrial FBC boilers and conventional industrial boilers with stack-gas cleaning.

Projected costs are largely uncertain; thus, direct comparisons between industrial FBC boiler costs and conventional boiler costs provide only rough perspective. The trends indicated by the costs, however, are probably valid.

Four industrial FBC boiler capacities are considered, ranging from 8.8 to 58.6 MW_t coal input. Three representative coals are considered, a high-sulfur bituminous coal, a low-sulfur bituminous coal, and a low-sulfur western subbituminous coal. Three general classes of sorbent reactivity (low, medium, and high) are defined and applied to the projections, with a single representative sorbent selected for each class.

Three sets of ACL for SO_x, NO_x, and particulate matter, as shown below, are applied as potential levels of control: stringent, intermediate, and moderate.

	SO_2	NO_x	Particulate Matter
Control Level	% Sulfur Removal	ng/J (lb/10^6 Btu)	ng/J (lb/10^6 Btu)
Stringent	90[a]	215 (0.5)	12.9 (0.03)
Intermediate	85[a]	258 (0.6)	43 (0.1)
Moderate	75[a]	301 (0.7)	107.5 (0.25)

[a] A ceiling level of 516 ng/J (1.2 lb/10^6 Btu) SO_2, and a floor level of 86 ng/J (0.2 lb/10^6 Btu) SO_2.

It is the impact of these three sets of ACL on the industrial FBC boiler performance and cost that is characterized in this study. A special case of "no sulfur removal required" is also discussed briefly.

The conclusions we have developed in this assessment follow.

IMPACT OF THE ALTERNATIVE CONTROL LEVELS

- FBC is competitive with conventional FGD, even for stringent levels of control. To assure competitiveness, however, FBC should be designed for adequate gas residence time in the bed. Increased capitalization cost resulting from increased gas residence time is more than offset by reduced costs resulting from reduced sorbent feed requirements. Not all commercial industrial FBC designs include gas residence times as long as the 0.67s considered here.

- The three control levels of ACL for sulfur dioxide (SO_2) considered significantly influence the required calcium-to-sulfur (Ca/S) feed ratio for the base boiler design, especially when coupled with the variation in sorbent activity. The moderate control level Ca/S ratio ranges from 1.9 to 6.4, the intermediate range from 2.4 to 7.4, and the stringent range from 2.4 to 8.5, for sorbents varying from high activity to low activity, respectively, with bituminous coal. The subbituminous coal requires less sorbent consumption because of its coal-ash alkali constituents.

- The variation in the three ACLs for SO_2 control have no significant influence on the mass rate of total waste solids produced by industrial FBC boilers at the base design conditions (<23% change). The sorbent activity has a greater impact on the mass rate of waste solids production for the high-sulfur bituminous coals (up to 77% change). For the low-sulfur coals the ACL has even less influence.

- The particulate control efficiency requirements for the final-stage control device are nearly independent of the sorbent sulfation activity and the level of the SO_2 control requirement.

- Conventional technology exists - electrostatic precipitators or fabric-filter techniques (baghouses) - that should be capable of controlling particulate emissions to the most stringent level of 12.9 ng/J (0.03 lb/MBtu). Testing these control options on operating fluidized-bed combustors is necessary, however, before actual control performance and any operating problems can be firmly defined. No such testing has yet been conducted.

- Fabric-filter techniques seem most appropriate for the industrial FBC boiler, with the major uncertainties in their use involving details of the equipment specifications (filter materials, face velocities, etc.), operating and maintenance procedures, and potential problems with carbon accumulation and baghouse fires.

- The ACL for NO_x control cause the greatest uncertainty. The moderate and intermediate control levels can be achieved with little or no modification to the base boiler operating conditions, but the stringent control level may require more sophisticated modifications, as, for example, two-stage combustion.

- The industrial FBC boiler energy conversion efficiency depends heavily on the coal sulfur content, sorbent reactivity, fluidized-bed operating conditions, and, to a lesser extent, the SO_2 emissions control standard. For a given set of fluidized-bed design and operating parameters, this effect is felt mostly through the sorbent calcination energy losses. If the SO_2 emission control level is changed from moderate to stringent, depending on the sorbent activity category and particle size, the boiler efficiencies for bituminous high-sulfur coal are reduced by 0.6 to 1.5 percent and, for the two low-sulfur coals, by 0.05 to 0.16 percent. This is less than the uncertainty in the combustion efficiency that will be achieved in the system with a given coal and design.

- The impact of the SO_2 control levels considered on the cost of the industrial FBC boiler is dependent on the coal and sorbent properties. The SO_2 emission control standards have little impact on the steam costs (less than 1 percent) for bituminous low-sulfur (0.9 wt % sulfur)

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and western subbituminous (0.6 wt % sulfur) coals. Depending on the sorbent consumption, however, the steam costs for bituminous high-sulfur coal (3.5 wt % sulfur) at the stringent control level (90% sulfur removal) are 4 to 10 percent higher than at the moderate control level (78.7% sulfur removal, 516 ng SO_2/J). For the three reference coals the steam conditions, saturated or superheated, have a greater effect on the steam costs than does the SO_2 control level. This difference is due to the increased capital investment for generating superheated steam and lower throughputs of superheated steam (kg steam/MW_t) than for saturated steam. If no sulfur removal is required, a significant cost reduction will be realized with the bituminous high-sulfur coal.

- The levels of particulate control will have little impact on the industrial FBC boiler cost, with similar particulate control equipment required to satisfy all levels of the standard. The NO_x ACL could affect the performance and cost of the industrial FBC boiler, especially at the stringent level, because of the possible need for modified operating conditions or equipment design.

COMPARISON WITH CONVENTIONAL INDUSTRIAL BOILERS

- The Ca/S molar ratio requirements projected for the base industrial FBC boiler in this study are at least two to three times those needed for lime/limestone wet-scrubbers (Ca/S $\cong$ 1.05 to 1.3) used for SO_2 control in conventional boiler systems for bituminous coals. For western subbituminous coals and lignites the Ca/S molar ratios for FBC may be comparable to those for lime/limestone wet-scrubbers because of the sulfur removal activity of the coal ash when burned at fluidized-bed boiler conditions.
- The production of solid waste for the industrial FBC boiler burning high-sulfur bituminous coal is from once to twice the amount of sludge (50% water) produced by a conventional boiler with lime/limestone wet-scrubbing. When low-sulfur coal is used, the waste solid from the industrial FBC boiler is comparable in production rate to sludge from a conventional industrial boiler. These wastes will probably be

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classified as nonhazardous under RCRA guidelines. Landfill seems to be the most promising option for disposing of FBC solid residue, with relatively less environmental impact than from the flue-gas desulfurization (FGD) sludges from conventional coal-fired plants.

- The industrial FBC boiler and plant efficiencies are higher than those of the uncontrolled conventional spreader-stoker units but lower than those of the uncontrolled pulverized coal-fired boiler. This conclusion is very sensitive to the carbon combustion efficiency achieved. The boiler efficiency of the industrial FBC boiler compares favorably with that of the controlled, conventional, industrial coal-fired boiler. The effect on energy usage for the control of particulate and NO_x for industrial FBC should be of a magnitude similar to that for conventional industrial coal-fired boiler systems.

- The capital investment and annual costs for the larger-capacity industrial FBC boiler plants and for conventional industrial coal-fired boiler plants with FGD systems are comparable. For a given plant size and fluidized-bed operating conditions, the annual costs depend strongly on the coal sulfur content and on coal and sorbent costs. The cost advantage for the industrial FBC boiler generally increases for low-sulfur coals.

- The levels of emission control considered do not significantly affect the competitive position of the industrial FBC boiler.

- The above conclusions are contingent upon the FBC unit being designed with sufficient gas residence time ($\gtrsim$0.67s) to minimize sorbent consumption. Some current, commercially proposed FBC designs do not meet this requirement.

DESIGN AND OPERATING PARAMETERS

The range of possible sorbent performance is very broad. Low-activity sorbent is characterized by performance that is very sensitive to design and operating conditions, especially bed particle size. The industrial FBC boiler designer can significantly influence the SO_2 control performance of the boiler by varying velocity and bed depth.

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Increased capital investment due to increased combustor size will normally be offset by reduced operating costs and improved performance. Industrial FBC boiler developers should be cautious in their use of such design options as above-bed feeding of coal and/or sorbent and recycle versus once-through operation of primary cyclone catch material. The resulting SO_2 control performance is intimately related to coal properties, attrition and elutriation phenomena, and the highly variable properties of available sorbents.

- As with sulfur removal control, the combustor design and operating conditions greatly affect the nature of particulate emissions. The combustor can be designed and operated (e.g., at low velocities) to limit particle attrition and elutriation, though rather efficient final-stage particulate control devices will still be required (87 to 99.5% for high-sulfur bituminous coal; 74 to 99% for the low-sulfur reference coals).
- Fluidized-bed design and operating variables such as the primary cyclone fines recycle option, the number of coal feed points and their location, bed temperature, bed depth, superficial velocity, excess air, and sorbent particle size affect the annual costs per energy output (e.g., steam cost) more than the total capital investment. These variables influence the combustion efficiency, sorbent calcination energy requirements, stack losses, and auxiliary power consumption, thus affecting the boiler as well as the plant efficiencies.
- Sorbent particle attrition may represent the greatest uncertainty and variability in the projection of industrial FBC boiler performance. While the key factors of average bed particle size and elutriation rate are directly related to sorbent attrition behavior, SO_2 control performance in all likelihood is not adversely affected by sorbent attrition.
- The major uncertainty with the use of primary cyclone recycle is that the high dust loadings expected in the boiler convective zone can result in erosion of the convective heat transfer surface. Placement of the convective surface following the primary cyclone may be required.

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- The operating conditions of the shop-fabricated industrial FBC boiler are less flexible than are those of field-erected electric utility FBC boilers, because of size limitations. The criteria for combustion efficiency and auxiliary power consumption applied in the case of utility FBC boilers, however, may be relaxed for the industrial FBC boiler.

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3. SCOPE OF THE STUDY

The object of this report is to assess, in general terms, what impact various levels of emission control for industrial coal-fired steam generators will have on the performance and cost of industrial fluidized-bed combustion. This study was conducted in order to provide background information for the EPA effort to develop New Source Performance Standards for air emissions from industrial boilers. The factors to be considered in order to provide general perspective on this issue are:

- The industrial FBC application, capacity, design, and operating conditions
- The coal properties
- The sulfur sorbent properties.

Only atmospheric-pressure industrial fluidized-bed combustors for steam generation application have been considered. These are referred to specifically in the body of this report as industrial FBC boilers and AFBC. Pressurized FBC applications (PFBC), cogeneration, air-heater, or process stream heater applications are not directly addressed. Steam applications ranging from indirect heating to consumption in chemical processes are considered.

Four industrial FBC boiler capacities are considered, ranging in thermal input from 8.8 to 58.6 MW_t. Representative base steam conditions are selected for each plant size.

Base design and operating conditions were selected for the industrial FBC boiler units on the philosophical basis of maximizing shop fabrication. The selected base design and operating conditions are representative of the current technological capabilities of FBC, though not necessarily representative of industrial FBC boilers being marketed at the present time. Commercially available industrial FBC boiler designs are discussed in Appendix A. In addition to the base design and operating conditions, a number of key design and operating parameters are examined parametrically.

Three representative coals having the properties shown in Table 1 have been evaluated.

Table 1

REFERENCE COAL COMPOSITIONS

Coal	Moisture	C	H	S	O	N	Ash	J/g (Btu/lb)
Eastern High-Sulfur Bituminous	8.79	64.80	4.43	3.50	6.56	1.30	10.58	27,450 (11,800)
Eastern Low-Sulfur Bituminous	2.83	78.75	4.71	0.90	4.91	1.50	6.90	32,100 (13,800)
Western Low-Sulfur Subbituminous	20.80	57.60	3.20	0.60	11.20	1.20	5.40	22,330 (9,600)

line table

These compositions are averages developed from coals listed in Babcock & Wilcox "Useful Tables for Engineers & Steam Users," 12th ed., 1972, and provided to Westinghouse by EPA for use in this study.

Three representative limestones (Table 2) have been selected for the technical/economic assessment. Thermogravimetric (TG) activity curves (at standard conditions) for these sorbents are shown in Figures 1 through 3. These sorbents fall into the general categories of

Table 2

REFERENCE SORBENTS SELECTED FOR STUDY

Sorbent Reactivity Category	Low	Medium	High
Reference Sorbent	Menlo[a]	Bussen[b]	Western 90% CaL[c]
Ca as $CaCO_3$, wt %	92.6	76.8	91.5
Mg as $MgCO_3$	2.2	8.1	6.8
Al as Al_2O_3	5.6	5.4	0.40
Fe as Fe_2O_3	0.71	1.5	0.33
Si as SiO_2	3.8	9.4	1.6
Na as Na_2O	0.11	0.59	0.035
K as K_2O	0.43	1.1	0.004
Sum	105.8	102.9	100.7

[a] Schildberg-Menlo Quarry, Adair Co., IA.

[b] Bussen Quarry, St. Louis Co., MO.

[c] Western Materials Company, Orleans, IN.

high, medium, or low reactivity. The criterion for this grouping is that, for a set of standard test conditions (840°C, 0.5% SO_2, 15% CO_2, 4% O_2, 1000-1200 μm particle diameter, nonisothermal calcination at 10°C/min), the extent of sulfation of the sorbent (its utilization) at a reaction rate of 0.001 min^{-1} is greater than 35 percent, between 20 and 30 percent, or less than 15 percent, respectively. These categories were developed from an extensive TG data base and FBC development unit data evaluations. The reference sorbents are representative of each activity category. General ranges of attrition behavior for the sorbents are also considered.

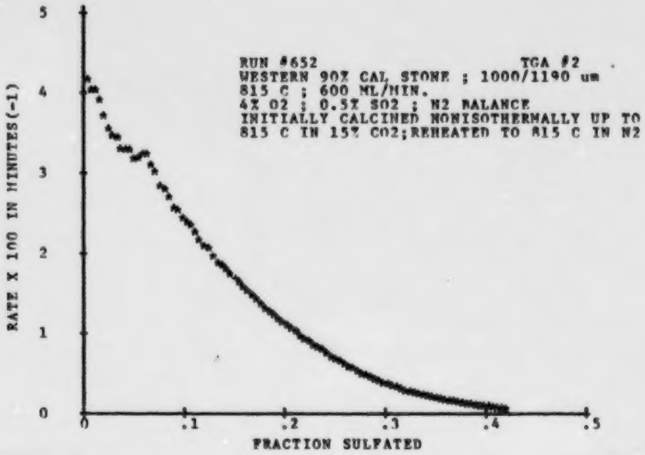

Figure 1 - High-Reactivity Sorbent TG Curve

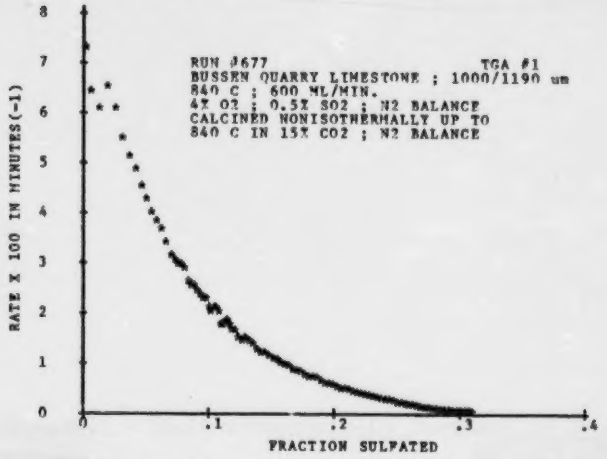

Figure 2 - Medium Reactivity Sorbent TG Curve

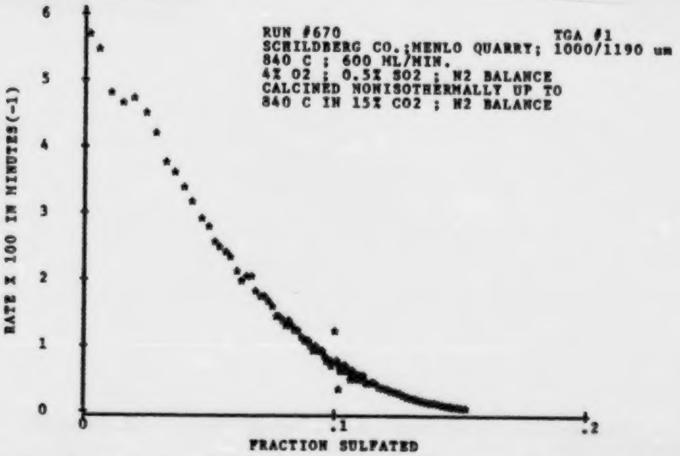

Figure 3 - Low-Reactivity Sorbent TG Curve

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While dolomites also have been found to be acceptable sorbents for AFBC, having consumption rates similar to those of limestones on a weight basis, only limestones are considered in this study.

Sink page

The pollutants currently being considered for regulation by NSPS in industrial boilers are SO_2, NO_x and particulate matter. Three potential levels of control for these three pollutants are considered in this study: stringent, intermediate, and moderate. A special, extreme case of zero sulfur removal required is also discussed. Table 3 presents the control requirements for each of these levels.

Table 3

POTENTIAL LEVELS OF CONTROL FOR INDUSTRIAL FLUIDIZED-BED BOILER SYSTEMS

use heads only

	SO_2	NO_x	Particulate Matter
Control Level	% Sulfur Removal	ng/J (lb/10^6 Btu)	ng/J (lb/10^6 Btu)
Stringent	90[a]	215 (0.5)	12.9 (0.03)
Intermediate	85[a]	258 (0.6)	43 (0.1)
Moderate	75[a]	301 (0.7)	107.5 (0.25)

[a] A ceiling level of 516 ng/J (1.2 lb/10^6 Btu) SO_2, and a floor level of 86 ng/J (0.2 lb/10^6 Btu) SO_2.

Figure 4 illustrates the three levels of control for SO_2 emissions. The required sulfur removal efficiency is plotted against the coal sulfur content, expressed as g sulfur/10^6 J of coal energy. Curves representing both the ceiling and floor levels are shown. The resulting levels of SO_2 control for the three reference coals are:

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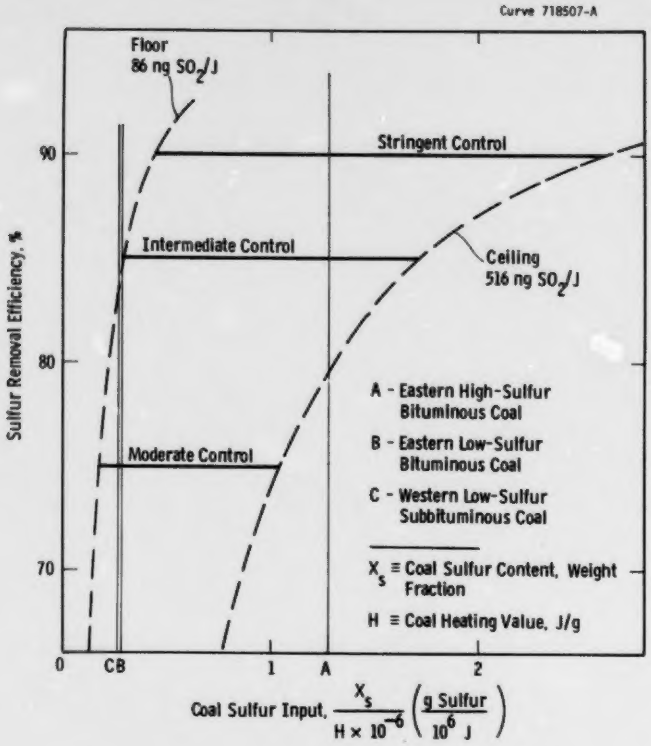

Figure 4 - Potential Levels of Control for SO_2

- Eastern high-sulfur bituminous coal: stringent - 90 percent sulfur removal; intermediate - 85 percent; moderate - 78.7 percent
- Eastern low-sulfur bituminous coal: stringent - 84.7 percent; intermediate - 84.7 percent; moderate - 75 percent
- Western low-sulfur subbituminous coal: stringent - 84.0 percent; intermediate - 84.0 percent; moderate - 75 percent.

The industrial FBC boiler plant performance (required sorbent feed rate, waste solids produced, general ability to achieve the ACL, overall plant energy efficiency) and cost (capital investment and steam cost) is projected for each alternative level of control as a function of the study parameters listed. Conclusions about preferred design and operating philosophy have been drawn.

In this study we have developed a base conceptual design of an industrial FBC boiler plant. The base design is used for the assessment of cost and performance, and a variety of design and operating parameters have been considered to test the cost and performance sensitivity. Within the expected accuracy of the projected costs, the results should be applicable to the broad range of industrial FBC boiler plants being currently proposed. That is, the industrial FBC boiler costs are much more sensitive to factors that are essentially constant for systems having significantly different design details, such as the amount of heat transfer surface, solids handling system capacity, and air compression requirements, than they are to the apparent design differences in the units, such as vessel shape and arrangement of functions.

The sensitivity of the costs and conclusions to the economic premises selected for the costing basis is also discussed.

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4. INDUSTRIAL AFBC CONCEPTS

Industrial AFBC concepts may be categorized in several different ways. The fluidization regime, which characterizes the gas-solid dynamics, is a common, fundamental basis for categorizing fluidization concepts. Major regimes identified are the bubbling, turbulent, fast, and entrained regimes.[1,2] Many subregimes are also defined in the literature; for example, slow bubbles, fast bubbles, exploding bubbles, shallow beds, and slugging beds.[3] The bubbling and turbulent regimes are classified as dense-bed regimes because of the dense packing of the particles in these regimes relative to that of the dilute-bed regimes, fast and entrained fluidization.

Figure 5 illustrates a qualitative mapping of the occurrence of these fluidization regimes. The region representing the AFBC dense-bed operation is identified in the figure. The regime selected for the fluidized combustor operation is of critical importance to its design and performance.

AFBC concepts may also be categorized by their method of bed temperature maintenance -- primarily, heat transfer surface within the bed heating a working or process fluid, excess air control, or the circulation of a stream of solids with intermediate cooling.

The AFBC concept can also be characterized by methods of maintaining carbon combustion at an acceptable level: the carbon burnup bed module, the recycle of elutriated carbon-bearing particles, or once-through carbon flow (with excess air, bed temperature, bed depth, bed velocity, and coal particle feed size distribution properly selected).

The general flow of solids in the AFBC system represents another basis for concept classification. In general, the dense-bed regimes result in significant streams of solids being both elutriated with the

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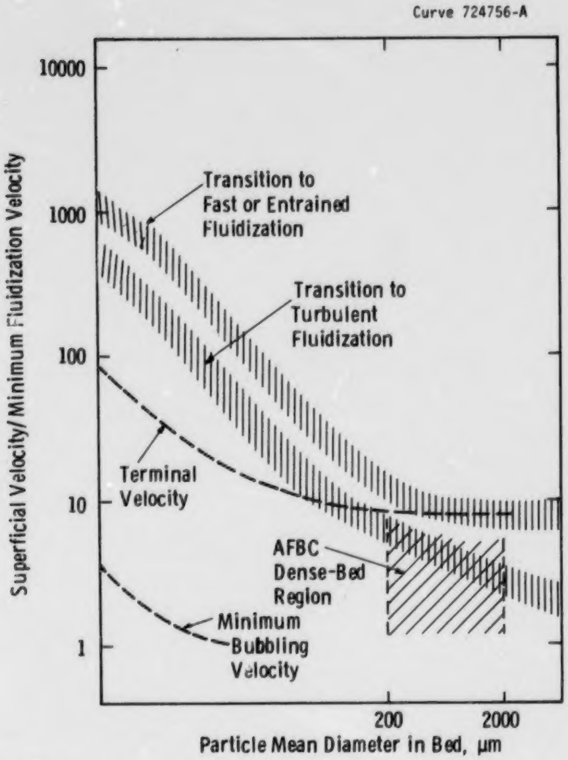

Figure 5 - Qualitative Characterization of Fluidized-Bed Combustion Regimes

combustion products and overflowing the bed (to maintain constant bed depth). The elutriate stream of solids may be collected in particulate control devices and recycled to the bed or may be stored for disposal (once-through elutriate). In the higher velocity, dilute-bed concepts all of the bed material (other than oversized material) generally is elutriated, and a substantial portion of the elutriate is recycled, the dynamics of the system being highly dependent on the recycle rate.

Other factors, such as the separation of major unit functions (combustion, sulfur capture, heat transfer) into discrete steps in series versus integrated operation; internal bed circulation pattern versus normal, uniform mixing; or modular/staging versus single-bed operation, are also common concept classes.

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Specific examples of available or proposed industrial AFBC concepts are listed below:

- Dense-bed regime concepts
  - The Georgetown University unit (Foster-Wheeler [FW], Pope, Evans and Robbins [PER]) under U.S. DOE sponsorship
  - The Rivesville unit (FW, PER) under U.S. DOE sponsorship
  - The Central Soya unit (Johnstown Boiler Co.)
  - The Central Ohio Psychiatric Hospital unit (Babcock Contractors, Inc.) under Ohio-DOE sponsorship
  - The Electric Power Research Institute (EPRI) test facility (Babcock & Wilcox [B&W])
  - The Tennessee Valley Authority (TVA) 20 MW_e pilot plant (B&W)
  - The Oak Ridge National Laboratory (ORNL) CCC unit (Westinghouse, B&W, Stone & Webster Engineering Corporation [SWEC])
  - Great Lakes Naval Training Center Unit (Combustion Engineering, Inc.) under U.S. DOE sponsorship
  - Wilkes-Barre anthracite culm unit (FW, PER) under U.S. DOE sponsorship
  - Shamokin Area Industrial Corporation anthracite culm unit (Curtiss-Wright, Dorr-Oliver, SWEC)
- Dilute-bed regime concepts
  - The Lurgi fast fluidized bed
  - The Ahlstrom fast fluidized bed
  - The Battelle Multisolid fluidized bed (under license to Struthers Wells Corp.)
  - The Westinghouse circulating bed (draft tube) concept.

In this report we consider only the dense-bed AFBC concept class, which represents the most highly developed and currently available technology for industrial AFBC. Dilute-bed concepts offer attractive characteristics and should also be considered for selected applications. Similar analyses that are applicable to dilute-bed concepts can be developed to provide perspective on the economic sensitivity original not provided

to the selection of design and operating conditions for achieving environmental standards. Note, however, that conclusions from the dense-bed analysis presented here are not directly applicable to the dilute-bed concepts.

5. BASE INDUSTRIAL FBC BOILER DESIGN SPECIFICATIONS AND STUDY PARAMETERS

The base industrial FBC boiler system design specifications are summarized here. A more detailed design description of the base conceptual design developed is presented in Appendix B. The ranges of design and operating parameters considered are also described.

The design philosophy we have applied in this study has been the maximization of shop fabrication and the standardization of fluidized-bed boilers. The maximum shippable boiler dimensions given in Table 4 have been used as guidelines.

Table 4

MAXIMUM SHIPPABLE BOILER DIMENSIONS[4]

Use head only

Location	Constraints*
East of Harrisburg, PA and north of Baltimore, MD	12 ft, 4 in wide and 13 ft, 10 in high or 10 ft, 6 in wide for the first 8 in of height and then 12 ft, 4 in wide with a total height of 14 ft, 4 in
West and southwest of Harrisburg, PA	12 ft, 6 in wide and 15 ft, 5 in high

*Maximum length of 60 ft.

Table 5

BASE AND PARAMETRIC INDUSTRIAL FBC BOILER OPERATING CONDITIONS

Condition	Base	Parametric
Bed Temperature, °C (°F)	843 (1550)	a
Superficial Velocity,[b] m/s (ft/s)	1.83 (6)	1.22 - 3.05 (4-10)
Expanded Bed Depth, m (ft)	1.22 (4)	Up to 2.44 (8)
Excess Air, %	20	Up to 100
Coal Feed Size Distribution, μm (in)	< 6350 (0.25)	a
Sorbent Feed Size Distribution, μm	< 2675 (0.125)	a
Bed Particle Size Distribution,[c] μm	500 and 1000	---
Bed Turndown Method	Velocity, bed temperature, module shutdown, and localized bed-section slumping	---

[a] Comments on the qualitative impact are presented.

[b] Base gas residence of 0.67s.

[c] Surface mean diameter of the bed sorbent.

The base operating conditions for the industrial FBC boiler (summarized in Table 5) have been developed from available data and technical assessments. These base conditions provide reasonable FBC performance [gas residence time (see Figures 8-12) and sorbent particle size for achieving acceptable SO_2 capture; acceptable unburned carbon losses, and sorbent elutriation losses[5]] in most circumstances. For example, a gas residence time of 0.67s was selected to promote minimum sorbent consumption.

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In addition, we have performed parametric trade-off studies to assess the impact of conditions even more favorable to FBC performance. These ranges of parametric operating conditions are also indicated in Table 5. The base boiler design specifications for the four plant sizes evaluated are given in Table 6. To assess the economic impact of the ACL in Table 3 for industrial coal-fired FBC boiler systems, conceptual designs have been prepared for the four plant capacities (a description is given in Appendix B). Several design alternatives have also been studied; they are listed in Table 7.

The industrial FBC boiler plants have boundaries defined at the inlets to coal and sorbent storage, to the economizer, and at outlets from the stack, spent solids on-site storage, and steam generators, excluding the steam lines to the point of utilization (see Figure 6). The coal and sorbent handling systems included are independent of each other and consist of receiving hoppers, silos, feed elevators, sizing and drying systems, screens, bin vibrators, surge bins, and necessary valves and piping. The spent solids handling system consists of surge bins, solids cooler, and a pneumatic transport system. The flue gas and combustion air system includes forced draft fans and drive motors, bed cyclone units, surge bins at dust coolers, a final particulate control device (fabric filters), a stack, and the necessary valves and ducting arrangement. The fluidized-bed combustor module includes the rectangular vessel (FBC cell), heat transfer surface (waterwall tubes and tube bundles), various inlet and outlet headers, a steam drum, water circulation pumps (forced), bed air dampers, fluidization grilles, ignition nozzles, coal and sorbent feeders, injection nozzles for recycle material, flues and ducts, and the associated insulation, piping, instrumentation, and boiler internals supports. The main steam system in an industrial FBC boiler plant will be similar to that in a conventional boiler plant and will include the boiler feedwater treating system, condensate system with appropriate piping, instrumentation, chemical feed, and pumping facilities.

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Table 6

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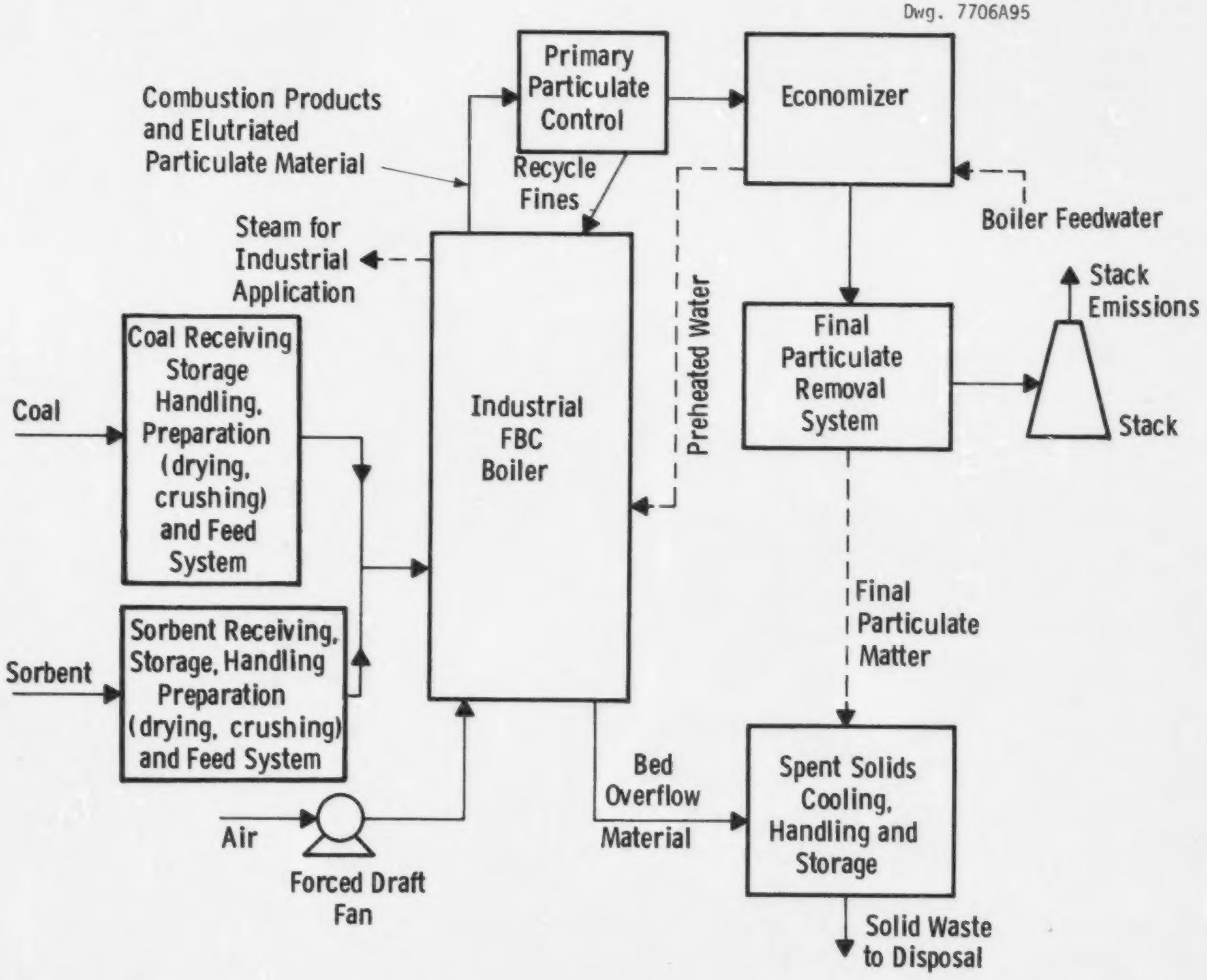
BASE DESIGN SPECIFICATIONS FOR THE INDUSTRIAL FBC BOILER

Thermal Input, MW_t (10^6 Btu/hr)	8.8 (30)	22.0 (75)	44.0 (150)	58.6 (200)
Steam Conditions/Feedwater Temperature, °C	166 kPa, sat'd/93	166 kPa, sat'd/93	469 kPa, 316°C/93	772 kPa, 399°C/93
Steam Water Circuitry	Natural plus forced circulation			
Number of Modules	1	1	2	3
Single Module Details:				
Number of beds/module	2	2	2	2
Zones in air plenum (for local slumping)	2	4	4	4
Bed cross-sectional area, m^2 (ft^2)	6.5 (70)	16.6 (179)	16.6 (179)	14.8 (159)
Overall boiler dimensions, l × w × h, m	3.96 × 3.05 × 4.57	9.75 × 3.05 × 4.57	9.75 × 3.05 × 4.57	8.84 × 3.05 × 4.57
Heat transfer surface in bed	0.254 m (1 in) dia. horizontal tubes with 0.183 m × 0.91 m (6 in × 3 in) diamond spacing			
Heat Transfer Coefficient, $W/m^2 - °C$ (Btu/hr, ft^2 °F)				
Tubes in bed	284 (50)	284	284	284
Tubes in convective pass	85 (15)	85	85	85
Tubes in economizer	40 (7)	40	40	40
Heat Transfer Surface Configuration:				
Super heating	None	None	Bed	Bed
Evaporation	Bed/conv. pass	Bed/conv. pass	Bed/conv. pass	Bed/conv. pass
Preheating	Economizer	Economizer	Economizer	Economizer
Coal and Sorbent Feed (premixed):				
Location	In bed	In bed	In bed	In bed
Distribution, m^2/feed point (ft^2)	0.93 (10)	0.93 (10)	0.93 (10)	0.93 (10)
Overflow Material Drain Points/Bed	1	1	1	1
Control of Carbon Utilization Technique	Reinjection of carry-over fines to main bed from primary cyclone			
Stack-Gas Temperature, °C	135	135	135	135
Spent Solids Cooling	With recovery of energy			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Combustion Products and Elutriated Particulate Material
Primary Particulate Control
Recycle Fines
Economizer
Boiler Feedwater
Steam for Industrial Application
Preheated Water
Industrial FBC Boiler
Coal
Coal Receiving Storage Handling, Preparation (drying, crushing) and Feed System
Sorbent
Sorbent Receiving, Storage, Handling Preparation (drying, crushing) and Feed System
Air
Forced Draft Fan
Final Particulate Removal System
Stack Emissions
Stack
Final Particulate Matter
Bed Overflow Material
Spent Solids Cooling, Handling and Storage
Solid Waste to Disposal

Figure 6 - Industrial FBC Boiler System Boundaries

Table 7

INDUSTRIAL FBC BOILER DESIGN ALTERNATIVES

Steam Conditions[a]	1135 kPa (150 psig), saturated to 5272 kPa (750 psig), 399°C (750°F)
Feedwater Temperature,[a] °C	27 to saturation
Heat Transfer Surface Configuration	Superheat in convective pass
Coal and Sorbent Feed	
Location	Above bed
Distribution, m^2 (ft^2)/Feed Point	Up to 3.72 (40) for in-bed feeding
Waste Solids Bed Overflow Cooling	Cooling without energy recovery
Flow Configuration for Primary Cyclone Catch	Nonrecycle of primary cyclone catch
Plant Equipment Simplification	Elimination of coal and sorbent crushing, drying, waste solids heat recovery

5272

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[a] Specifications will depend on particular application.

6. SULFUR DIOXIDE EMISSION CONTROL

The sorbent consumption rate required to meet the ACL for SO_2 emissions is of key importance to the overall performance (energy efficiency, environmental impact, cost) of the industrial FBC boiler. In this section the sulfur removal performance of the base boiler is presented, along with a discussion of the effects of several design and operating parameters.

Sulfur dioxide emission control performance for the industrial FBC boiler has been projected using semiempirical models developed by Westinghouse. These models have been previously applied to the evaluation of electric-utility-size FBC boilers.[5] The procedure for projecting required Ca/S ratios is described in Appendix C.

When fed to the FBC boiler, the sulfur in the coal is released primarily as SO_2, which reacts with the calcined sorbent in the FBC bed to form calcium sulfate ($CaSO_4$). In essence, desulfurization in the bed is a competition between two processes: once SO_2 is released in the bed, it may either remain in the gas phase and escape with the products of combustion, or it may react with calcium oxide (CaO) in the bed to form $CaSO_4$. The balance between these processes can be drastically altered by enhancing the probability of escape (shallow bed, high gas velocity, SO_2 release near the bed surface, above-bed SO_2 release, nonuniform coal feed distribution), or by increasing the reaction rate of the sorbent (e.g., decreasing the sorbent particle size in the bed, selecting the optimum bed temperature, selecting a more reactive sorbent). The sulfation kinetics of limestones and dolomites can vary greatly from one quarry source to another, and pore diffusional limitations, known to be a function of both the natural sorbent properties and the calcination conditions, have been observed to play a key role in sorbent utilization.[6]

The model that Westinghouse developed incorporates the following basic assumptions:

- Release of sulfur from coal as SO_2 due to char and volatile combustion occurs uniformly throughout the combustor bed of AFBC units.
- The rate-limiting process for SO_2 capture in the bed is governed by diffusion within the sorbent particle itself.
- Sorbent reactivity is a function of the bed calcining conditions and the degree of sulfation and is not independently affected by the residence time of sorbent particles in the bed.

In making the sorbent consumption projections, we have accounted for factors such as the volume fraction of bed bubbles, bed voidage in the emulsion phase, the fraction of the emulsion volume occupied by inerts, and the fraction of the bed volume occupied by the heat transfer surface. Credit has been taken for coal-ash alkali sulfur capture with the subbituminous coal, since this contribution can be extensive.[7]

BASE INDUSTRIAL FBC BOILER SULFUR REMOVAL PERFORMANCE

The influence of sorbent type and the average sorbent particle size in the bed on the Ca/S molar feed ratio needed for limiting SO_2 emissions from a FBC boiler operated at the base conditions with bituminous coals is shown in Figure 7, as predicted,using the Westinghouse model. The Ca/S molar ratio requirements for the high-reactivity category, medium reactivity, and low reactivity sorbents using 1000 μm average sorbent particle diameter for achieving 75 to 90 percent sulfur removal efficiency in the bed varies between 2.3 and 3.5, 2.9 and 4.3, and 6.0 and 8.5, respectively. The particle size represents the surface mean diameter of particles in the bed. The projected Ca/S ratios for the three reference coals are listed in Table 8. The sorbent mass feed rate requirements for the three reference coals to meet the three SO_2 emission control levels are summarized in Table 9. Figure 4 illustrates the way in which these three levels of SO_2 emission control are determined, based

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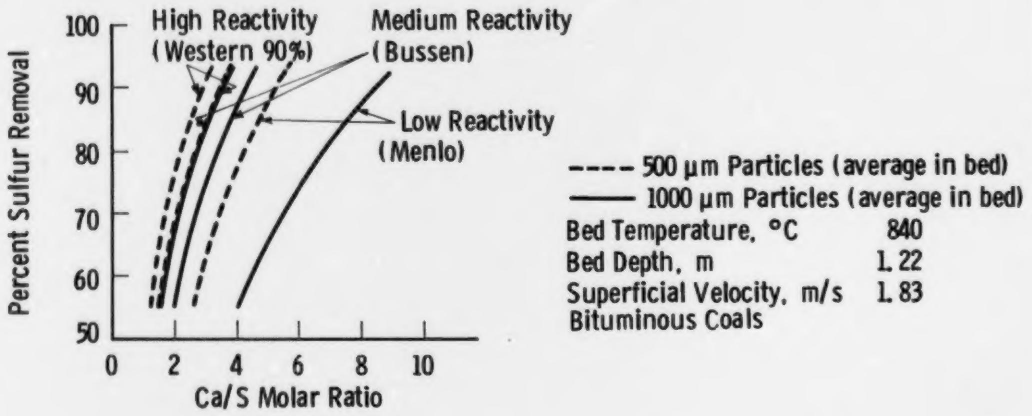

Figure 7 - Effect of Sorbent Activity Category and Particle Size on Ca/S Molar Feed Ratio Requirement for the Base Operating Conditions

on the floor and ceiling values and the sulfur removal efficiency requirement. The amounts of low-reactivity-sorbent needed are considerably higher than those for the other two sorbent activity categories, and the results also suggest the desirability of operating with beds having average particle diameters as small as can be maintained to minimize sorbent consumption. An average particle diameter of 500 μm is probably a minimum for representative operating conditions for dense-bed combustors.

Table 8 Dwg

Table 8

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REQUIRED Ca/S MOLAR RATIOS

Sorbent Activity Category	High		Medium		Low	
Average Bed Particle Diameter (surface mean), μm	500	1000	500	1000	500	1000
SO_2 Emissions Control Standard (% Sulfur Removal)						
	Bituminous High-Sulfur Coal (3.5 wt% S)					
Stringent (90)	2.8	3.5	3.4	4.3	5.3	8.5
Intermediate (85)	2.5	2.9	2.9	3.7	4.7	7.4
Moderate (78.7)	2.1	2.5	2.5	3.1	4.1	6.4
	Bituminous Low-Sulfur Coal (0.9 wt % S)					
Stringent & intermediate (84.7)	2.4	2.8	2.9	3.6	4.6	7.2
Moderate (75)	1.9	2.3	2.3	2.9	3.9	6.0
	Western Subbituminous Coal (0.6 wt % S)					
Stringent & intermediate (84.0)	1.1	1.3	1.3	1.7	2.4	3.2
Moderate (75)	0.7	0.9	1.0	1.2	1.8	2.1

Table 9

Dwg. 7737A12

SORBENT REQUIREMENTS

Sorbent Activity Category	Sorbent Consumption, Mass/Mass of Coal					
	High		Medium		Low	
Average Bed Particle Diameter (surface mean), μ.	500	1000	500	1000	500	1000
Emission Standard (% Sulfur Removal Required)						
	Bituminous Coal (3.5 wt% S; Coal Rate = 36.43 μg/J)					
Stringent (90)	0.34	0.41	0.49	0.62	0.62	1.00
Intermediate (85)	0.30	0.35	0.42	0.52	0.55	0.87
Moderate (78.7)	0.25	0.29	0.36	0.45	0.49	0.76
	Bituminous Coal (0.9 wt% S; Coal Rate = 31.15 μg/J)					
Stringent & Intermediate (84.7)	0.07	0.09	0.10	0.13	0.14	0.22
Moderate (75)	0.06	0.07	0.09	0.11	0.12	0.18
	Subbituminous Coal (0.6 wt% S; Coal Rate = 44.78 μg/J)					
Stringent & Intermediate (84)	0.02	0.03	0.03	0.04	0.05	0.07
Moderate (75)	0.01	0.02	0.02	0.03	0.04	0.04

EFFECT OF COMBUSTOR DESIGN OPTIONS AND OPERATING PARAMETERS

The major design options and operating parameters that apply to the industrial FBC boiler base system evaluated above are:

- Coal feed point number and location
- Nonrecycle operation of the primary cyclone catch
- Superficial velocity
- Bed depth
- Bed temperature
- Excess air
- Coal feed size distribution.

The effect of these parameters on SO_2 control is briefly discussed here, along with comments on some major performance assumptions:

- Fraction of inerts in the bed
- Fraction of the coal sulfur released
- Sulfur removal activity of subbituminous coal ash.

The fluidized-bed design variables that can affect the SO_2 generation pattern in the bed are:

- The location of coal feed points and their number
- The feed coal particle size distribution
- The extent of vertical solids mixing in bed (related to bed velocity, bed particle size, internals in bed, etc.)
- The excess air level.

The sulfur removal model form, for AFBC dense-bed application, is:

$$E_b = 1 - \left[\frac{U}{k'h} 1 - \exp\left\{-k'\frac{h}{U}\right\}\right] ,$$

where E_b is the sulfur removal efficiency of the unit (the fraction of the sulfur released from the coal that is captured by the bed, including the splash zone above the dense bed), U is the superficial velocity through the bed, h is the expanded bed depth, and k' is an empirical reaction rate constant given by:

$$k' \equiv k(1 - \delta)(1 - \varepsilon)(1 - f_I)(1 - f_H) ,$$

where f_H is the volume fraction of the total bed volume occupied by tube bundles and other internals, f_I is the volume fraction of the bed particle volume consisting of inert particles (coal char, coal ash, rocks), δ is the volume fraction of the bed occupied by bubbles, ε is the voidage of the bed emulsion phase [or the quantity $(1 - \delta)(1 - \varepsilon)$ is the volume fraction of particles in the bed], and k is the reaction rate constant determined by TG experiment. The rate constant k is a function of the temperature, pressure, excess air level, sorbent type, sorbent particle size (the surface mean diameter of particles in the bed), and the fraction of the calcium sulfated, α.

$$k \equiv \left(\frac{d\alpha}{dt}\right)_{TG} \frac{\rho}{C_{TG}} ,$$

where $(d\alpha/dt)_{TG}$ is the sulfation rate measured on the TG, ρ is the molar calcium density of the sorbent, and C_{TG} is the SO_2 concentration used in the TG experiment. The surface mean particle diameter in the bed must be estimated from combustor data or appropriate models. This aspect of the model is discussed in Section 8 (Particulate Control).

The overall sulfur removal retention of the unit, E, depends upon the fraction of the coal sulfur content release, η:

$$E = 1 - \eta(1 - E_b) ,$$

where the fraction of sulfur released may be less than unity because of incomplete combustion and retention of sulfur by coal ash constituents.

The molar feed ratio of calcium to sulfur is then:

$$Ca/S = (E + \eta - 1)/\alpha ,$$

where α represents the average sulfation level of sorbent particles in the bed.

The Ca/S ratio required to meet the alternative control level for SO_x is illustrated in the following five figures based on the relationships presented. The first three figures represent the bituminous coal case, the last two figures represent the subbituminous coal case. The gas residence time (h/U) is the key variable in these figures, with the sorbent reactivity classification and the surface mean particle diameter of the bed as parameters in the figures. The results are presented as narrow bands of Ca/S ratio for each set of conditions rather than by single curves. This band results from the potential range in bed expansion that could exist at any given value of the gas residence time because of the range of velocities that could apply to that time.

Parameter values estimated in evaluating the model relations and constructing Figures 8 through 12 are:

ρ = 3.5 x 10^{-2} moles calcium/cm^3

C_{TG} = 5.47 x 10^{-8} moles SO_2/cm^3

ϵ = 0.4

δ = 0.35 - 0.55 (over the potential range of operating velocities)

$f_H \simeq$ 0.05 (actually a function of the gas residence time and the number of modules)

$f_I \simeq$ 0.1 for bituminous coal

$\simeq$ 0.5 for subbituminous coal

$\eta \simeq$ 0.9 for bituminous coal

$\simeq$ 0.54 for subbituminous coal (if we assume 45% sulfation of the 20 wt % calcium oxide in the coal ash).

Previous studies have found these values to be representative of AFBC conditions.[11-15] The projected Ca/S ratios are also representative of test results from AFBC units.

Above-bed coal feeding or nonuniform in-bed feeding of coal can affect the SO_2 generation pattern in the bed and may result in increased SO_2 emissions because of inefficient gas-solid contacting, reduced SO_2 residence time in the bed, and increased SO_2 released above the bed. Attempts have been made to model the effect of above-bed feeding[8] and coal feed-point spacing,[9,10] but quantitative projections are not currently possible. Above-bed coal feeding requires the use of

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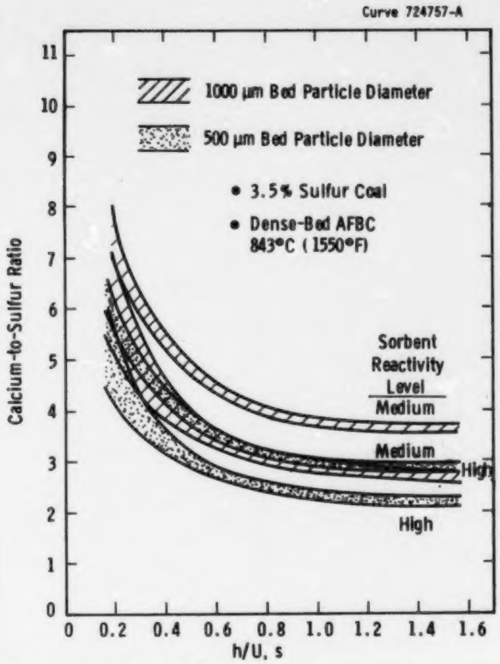

Figure 8 - Stringent ACL - 90% Sulfur Removal, Bituminous Coals

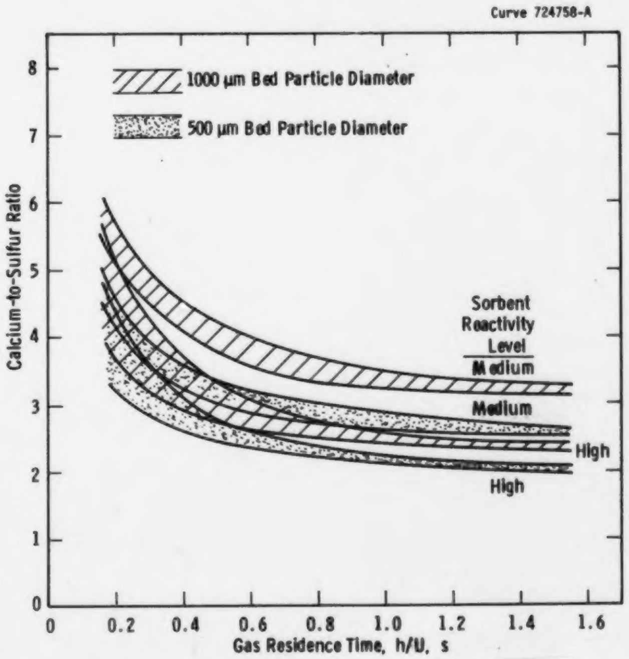

Figure 9 - Intermediate ACL - 85% Sulfur Removal, Bituminous Coals

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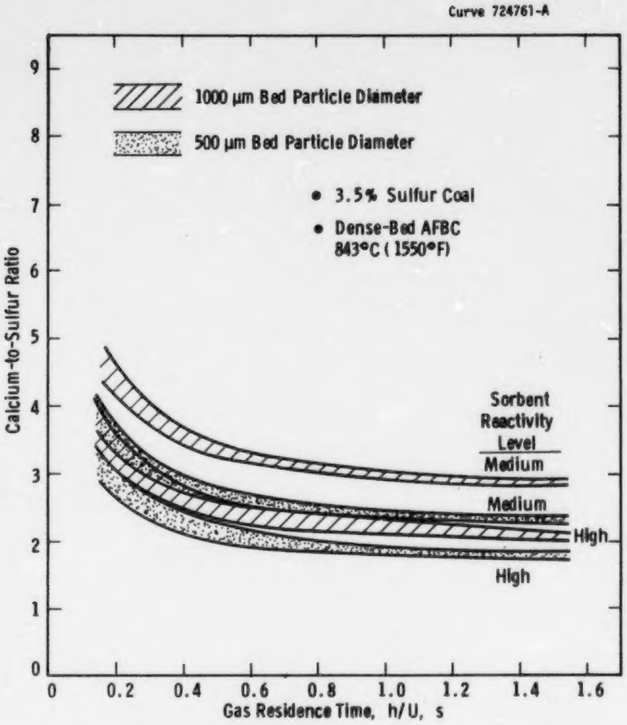

cve 724761-A

Figure 10 - Moderate ACL - 78.7% Sulfur Removal, Bituminous Coals

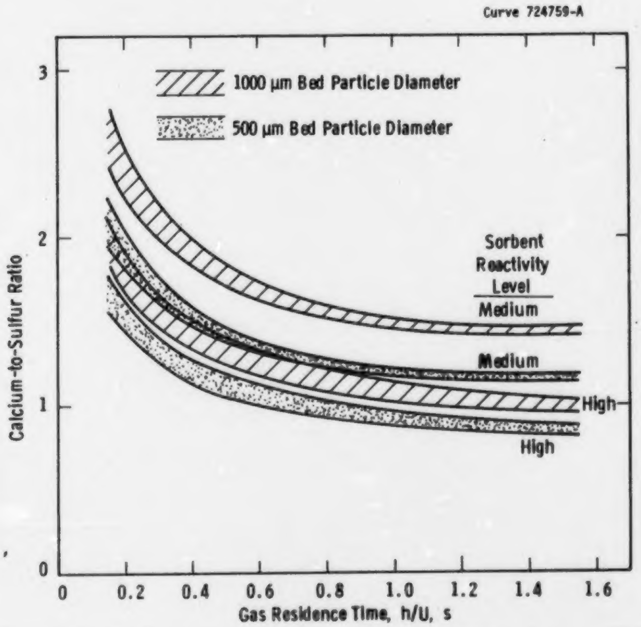

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Figure 11 - Stringent and Intermediate ACL - 84% Sulfur Removal, Subbituminous Coal

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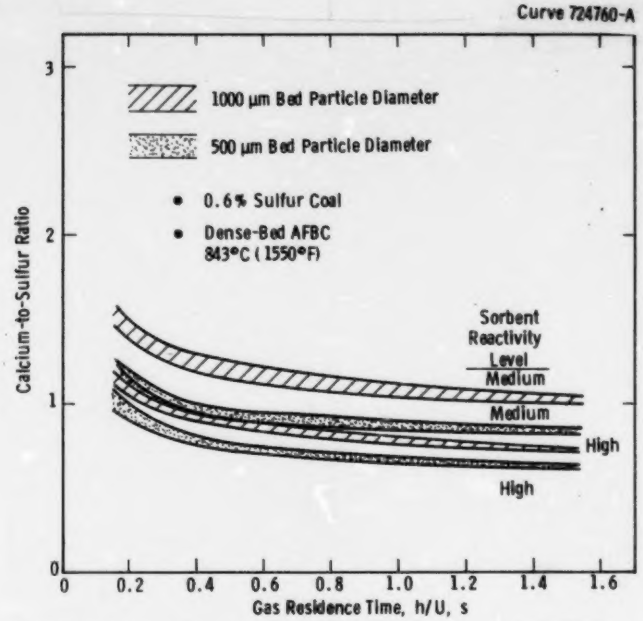

Figure 12 - Moderate ACL - 75% Sulfur Removal, Subbituminous Coal

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coal and sorbent carefully classified according to size, possibly with wet coal to reduce the rate of devolatilization. Industrial FBC boilers are being proposed and marketed with above-bed feed, but their performance has not yet been demonstrated (see Appendix A).

In the base design the elutriated particles collected by the primary cyclone are reinjected into the combustor bed for improved carbon and sorbent utilization. Sorbent particle attrition can result in increased concentration of fine particles in the bed by both coarse particle splitting and surface abrasion mechanisms and, hence, in increased elutriation. (Attrition is discussed further in Section 7.) The increased concentration of fine particles in the bed due to sorbent attrition and fine particle recycling can significantly reduce the surface mean particle size in the bed and possibly enhance sulfur removal efficiency. The recycled fine particles consist of unburned char and coal-ash and partially sulfated sorbent particles. The injection of the inert materials (coal ash) can dilute the concentration of sorbent particles in the bed and, hence, can also have an adverse effect.

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In the base design evaluation we have assumed a negligible inert particle volume fraction in the bed. The inert volume fraction will not have a significant effect until it exceeds about 20 percent.[16]

In the case of industrial FBC boiler operation without recycling the primary cyclone catch to the combustor bed, combustion efficiencies can be significantly less than those with the recycle option. Some of the sulfur in the coal feed, thus, will not be released and can escape along with the unburned char. The amount unreleased depends strongly on the distribution of sulfur between volatiles and char. Under such circumstances the sorbent consumption rate, being a direct function of the amount of sulfur released, will be reduced by at least the fraction of the sulfur retained in the char.

Coal feed size distribution and feed point location and distribution must be selected to prevent excessive combustion of char particles and coal volatiles in the freeboard, where sulfur capture is limited. The carbon burnup bed concept could potentially be detrimental to the system sulfur removal performance, as well as to NO_x and particulate control, depending on how the burnup bed is designed and operated. In general, use of the carbon burnup bed would not be recommended.

Other factors to be considered are the bed temperature, coal feed size distribution, coal feed point location, distribution of feed points, and the method of carbon combustion control. For bituminous coals the temperature range of 815 to 871°C (1500 to 1600°F) is generally appropriate for combustion and sulfur removal, but even this narrow range has a significant effect. For subbituminous coals and lignites lower bed temperatures (∿1300 to 1450°F) will be sufficient for high combustion efficiency (due to higher activity with these coals) and will maximize sulfur capture by the alkali constituents of the coal ash.[14,15]

With respect to turndown, fluidized-bed operation at partial load with less firing of fuel and, hence, lower gas flow rates may require lower Ca/S ratios to meet a given SO_2 emission control level. On the other hand, if used for load control, a decrease in bed depth can shorten the gas residence time in the bed and, thus, may require a greater Ca/S ratio for a given sulfur removal efficiency. The other parameter that may be adjusted for load control is the bed temperature, which will have a direct influence on the sulfation kinetics, the impact generally decreasing for smaller sorbent particle sizes. With modular combustors and sectional slumping of beds for turndown, the impact of turndown on SO_x control may be insignificant.

For subbituminous coals and lignites the coal ash typically contains significant proportions of alkali species (CaO, magnesium, oxide [MgO], sodium oxide [Na_2O], potassium oxide [K_2O], etc.) that can contribute to a reduction in the amount of sulfur released from the coal. Under fluidized-bed combustion operating conditions as much as about 50 percent of the coal sulfur has been found to be captured by the coal ash.[7] The resulting effect on the required Ca/S molar feed ratio is significant.[16]

CONCLUSIONS

The Ca/S molar ratio requirements projected for the base industrial FBC boiler with bituminous coals are at least two to three times those needed for lime/limestone wet-scrubbers (Ca/S $\cong$1.05 to 1.3) used for SO_2 control in conventional boiler systems using eastern coals. For western subbituminous coals and lignites the Ca/S molar ratios may be comparable to those for lime/limestone wet-scrubbers because of the sulfur removal activity of the coal ash at fluidized-bed boiler conditions.

The range of sorbent performance available is very broad. The low-activity sorbent category is characterized by performance much more sensitive to design and operating conditions, especially bed particle size, than high- and medium-activity sorbents. Operating parameters such as velocity and bed depth can be used by the industrial FBC boiler designer to influence the SO_2 control performance of the boiler. Design options such as above-bed feeding of coal and/or sorbent, and recycle or once-through operation of primary cyclone catch material, should be treated with caution by industrial FBC boiler designers, as the resulting SO_2 control performance is intimately related to the coal properties, attrition and elutriation phenomena, and the highly variable properties of available sorbents.

The ACL applied can significantly affect the sorbent consumption rate, but this impact will be no greater than that of the sorbent reactivity or the bed particle size parameters.

There is little benefit to reduced sorbent consumption from increasing the gas residence time above about 1 s (i.e., increased bed depth or reduced velocity). The minimum Ca/S ratio achievable will be about 2.0 for bituminous coals with a high-reactivity sorbent and fine particle size (with fine sorbent feed size distribution and/or high primary cyclone recycle rates to promote small surface mean sorbent particle diameter in the bed) and 0.7 for subbituminous coal with a high-reactivity sorbent and fine particle size.

In selecting the operating conditions for an industrial AFBC dense-bed boiler, one should minimize the sorbent consumption rate by maximizing the gas residence time and minimizing the bed particle size within the economic and boiler size constraints imposed on the system.

7. EFFECT OF SORBENT PARTICLE ATTRITION ON INDUSTRIAL FBC BOILERS

In FBC systems the attrition, or size reduction, of sorbent particles supplied to the system is of major significance to the performance of the system (see Section 6). In this section the current understanding of attrition in FBC systems is briefly reviewed and the impact of sorbent particle attrition on plant performance (SO_2 control, particulate control, solid waste, erosion, etc.) discussed.

SORBENT PARTICLE ATTRITION PHENOMENA

Sorbent particles supplied to an FBC system are exposed to a variety of environments, some in series and some in parallel, that can lead to their reduction in size and increase in number. Table 10 lists the major sources of particle attrition that have been identified.

The direct consequences of sorbent particle attrition are a reduction in the average sorbent particle diameter in the bed and an increase in the elutriation rate of sorbent from the bed relative to that that would prevail if no attrition occurred. These two quantities, the average bed particle diameter and the bed elutriation rate, are of great significance to the performance of the fluidized-bed combustion system.

Three sets of factors control the combustor attrition performance:

- The sorbent properties
- The combustor operating conditions
- The combustor system design.

While it is not currently possible to relate the FBC system attrition performance (i.e., the bed average particle size and particle elutriation

Table 10

SORBENT ATTRITION SOURCES IN FBC SYSTEMS

Sorbent Handling and Feed System

- Solids impaction in transfer lines, control valves, and injection nozzles

Fluid-Bed Combustor

- Grid jet attrition in the zone near the distributor grid
- Bed-mixing abrasion in the bubbling zone
- Splash region attrition above the bed
- Thermal shock on injecting cold sorbent feed and recycle fines to the combustor bed
- Internal gas pressure increase (CO_2 release due to calcination; evaporation of interstitial or occluded water)
- Structural changes due to calcination, sulfation, or hydration affecting particle strength

Particulate Control System

- Solids impaction in cyclones, transfer lines, reinjection system

---

rate) to these three controlling sets of factors, we can draw a number of conclusions from available FBC development unit data and from specific test programs addressing attrition.

WESTINGHOUSE ATTRITION MODELING CONCLUSIONS

Westinghouse has conducted a comprehensive program for EPA to investigate sorbent attrition mechanisms in FBC systems and their effect on fluidized-bed performance as a function of sorbent type and fluidized-bed combustor design and operating variables. Detailed results of this effort have been published.[11,17-19] We cannot at this time generalize the influence of various sources and mechanisms of sorbent attrition on fluidized-bed attrition performance; attrition tests simulating specific mechanisms need to be performed for each specific sorbent.

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In general, the sorbent attrition rate is a strong function of sorbent properties (crush strength, shear strength, etc., related to particle porosity, grain size, impurities, and so on).

Figure 13 shows the broad range of sorbent performance possible for a variety of limestones and dolomites tested under similar test conditions in hot, batch fluidization tests. Sorbent performance ranges from highly attrition resistant to highly attriting in nature. Quincy limestone is an example of a sorbent that explodes on initial heating at a relatively low temperature.

Similarly, the attrition performance of sorbents varies broadly over the range of FBC operating and design conditions. The dominant sources of particle attrition for a given sorbent cannot be predicted without specifying the FBC operating conditions and design details. It follows, also, that sorbent particle attrition can be minimized by proper selection of operating and design conditions.

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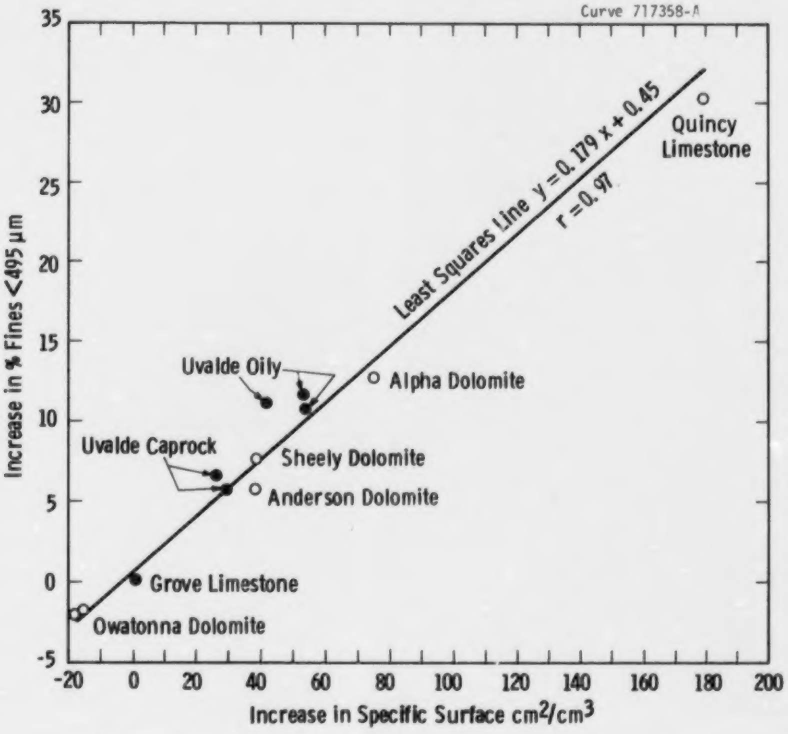

Figure 13 - Comparison of Sorbent Attrition Behavior

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Test data we have generated at Westinghouse clearly show that the sorbent attrition performance is a function of the sorbent history. For example, Figure 14 shows how the sorbent particle attrition rate changes with particle age and passes from an initial falling-rate period to a constant-rate period for the bubbling bed abrasion mechanism of attrition.[17] In general, the time-dependent nature of attrition must be accounted for in making performance projections.

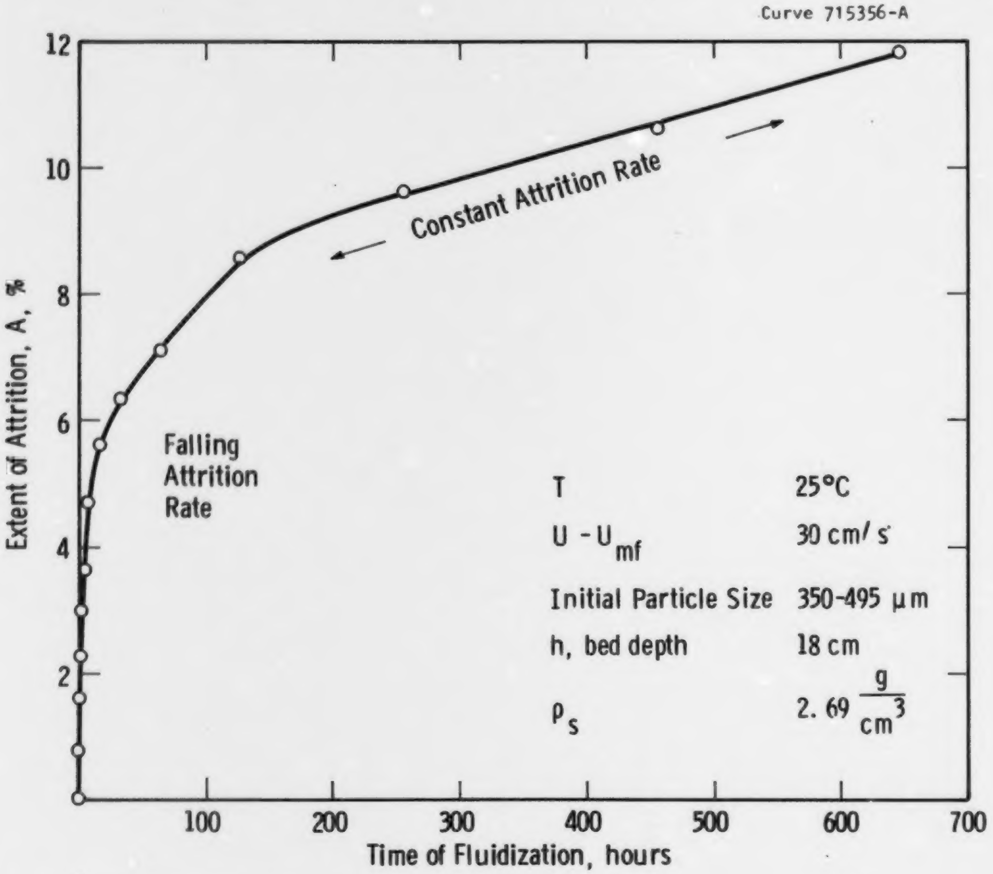

Figure 14 - Change of Extent of Attrition with Time in Bubbling Fluidization

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FBC DEVELOPMENT PLANT DATA ON ATTRITION PERFORMANCE

Numerous development units for FBC have been operated. The data generated have been reported in the literature. Only in a limited number of cases, however, are data collected and reported in sufficient detail to determine the extent of attrition occurring in the tests.

Extensive sorbent particle attrition and elutriation are common in FBC plant performance. Table 11 presents results deduced from data reported for the B&W 2 m by 2 m (6' x 6') unit[20] (under EPRI sponsorship) and the Morgantown Energy Technology Center (METC) 0.5 m diameter (18 in) unit,[21] both organizations reporting relatively complete data. Two bases for reporting attrition results are presented, the first as a percentage of the feed sorbent calcium rate and the second as a percentage of the bed calcium content per hour.

IMPACT OF SORBENT ATTRITION BEHAVIOR

The impact of sorbent attrition behavior on industrial FBC boiler performance (SO_2 control, particulate control, solid waste, erosion, etc.) is indeed significant.

Since many of the design and operating parameters that strongly influence sorbent particle attrition (for example, the combustor superficial velocity) also strongly influence the combustor performance (e.g., the gas residence time and the carbon combustion efficiency), minimized sorbent particle attrition behavior would probably not occur at conditions giving optimum overall combustor performance or cost. In fact, with respect to SO_2 control, much of the evidence suggests that sorbent particle attrition is not detrimental to the combustor performance, but results in improved SO_2 control performance by reducing the average bed particle size and increasing the content of highly active fine particles in the bed. Recent tests by FluiDyne using highly attriting sorbents, resulting in 100 percent elutriation of the feed calcium, achieved SO_2 control levels very similar to those that would have been achieved with attrition-resistant sorbent.[22]

Table 11

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FBC DEVELOPMENT UNIT DATA ON ATTRITION

B & W Run[14]	Sorbent	Feed Particle Size[a] μm	Bed Particle Size[a] μm	Velocity, m/s	Fraction Feed Calcium Elutriated	Percent Feed Calcium Attrited	Percent Bed Calcium Attrited/hr
1-1	Carbon Limestone	476	1082	2.7	0.333	19.1	2.53
1-2		947	1151	2.7	0.545	45.5	3.08
1-3		339	997	2.7	0.675	46.9	4.00
1-4		456	1206	2.2	0.494	46.5	2.28
1-5		433	1118	2.7	0.570	45.2	3.82
2-1		306	901	2.8	0.706	42.8	2.72
2-2		515	938	2.5	0.483	36.1	2.35
METC Run[15]							
PW-1, 1[b]	Greer Limestone	4288	1168	0.9	0.096	--	--
PW-1, 2[b]		894	1676	0.9	0.040	0	0
PW-1, 3[b]		894	1183	0.9	0.075	--	--
PW-1, 4[b]		4288	1569	1.1	0.184	19.6	0.43
PW-1, 5		4288	1062	1.1	0.145	14.5	0.51
PW-1, 6	Tymochtee Dolomite	550	216	1.0	0.453	25.6	0.66
PW-1, 7[b]		550	826	1.0	0.161	0	0
PW-1, 8		550	2303	1.1	0.811	57.2	1.00
PW-1, 9[b]		3548	747	1.1	0.355	37.5	1.08
PW-1, 10[b]		3548	701	1.2	0.259	28.8	0.80
PW-1, 11		3548	1474	1.2	0.288	28.8	0.80
PW-2, 1[b]	Greer Limestone	4288	845	1.1	0.279	29.0	1.25
PW-2, 2[b]		4288	574	1.0	0.438	47.4	1.67
PW-2, 3[b]		4288	513	1.0	0.271	35.4	0.96
PW-2, 4[b]		4288	431	1.1	0.262	38.6	1.71
PW-2, 4A[b]		4288	2269	1.1	0.609	61.2	2.25
PW-2, 5[b]		4288	1788	1.0	0.143	14.6	0.84
PW-2, 6[b]		4288	779	1.0	0.324	33.5	3.25
PW-2, 7		4288	1675	1.0	0.121	12.4	0.63

a - Surface mean particle diameter.

b - Primary cyclone catch recycled to combustor. for these runs the elutriation rate is defined as the rate of calcium passing through the primary cyclone.

The use of recycle from the primary cyclone, as has been assumed in the base industrial FBC boiler design, may result in increased concentrations of fine sorbent particles in the bed, thus effectively reducing the surface mean particle size in the bed, increasing the fine sorbent particle residence time in the bed, and reducing the sorbent consumption rate. The increased concentration of fine particles in the bed, however, can also result in increased particulate loading to the primary cyclone, consequently affecting the performance of the particulate emission control system.

Recycle of the primary cyclone catch to the combustor bed will significantly increase the dust loading of the combustion products passing through the convective heat transfer surface (by as much as an order of magnitude). The threat of erosion damage is high unless the appropriate design is utilized.

The particle attrition performance, also, will influence the nature of the waste solids issuing from the industrial FBC boiler plant. The size distribution, extent of calcination and sulfation, and point of collection of the waste solids are sensitive to attrition performance.

For low-sulfur coals, especially lignites and subbituminous coals, the relative rates of production of spent sorbent to coal ash, and elutriated fine sorbent particles resulting from sorbent attrition to coal ash, are small. Sorbent particle attrition will have less impact on all aspects of FBC boiler performance for the low-sulfur coals. Thus, sorbent selection criteria based on attrition properties of the sorbent can be relaxed for the low-sulfur coals.

CONCLUSIONS

Sorbent particle attrition is the greatest source of uncertainty and variability in projecting industrial FBC boiler performance. The key factors of average bed particle size and elutriation rate are directly related to sorbent attrition behavior. Attrition can range from 10 to 80 percent of the feed limestone rate, or 0.5 to 5 percent of the bed weight per hour.

Since many of the design and operating parameters that strongly influence sorbent particle attrition also strongly influence the combustor performance, minimized sorbent particle attrition behavior would probably not occur at conditions giving optimum overall combustor performance or cost. Much of the evidence suggests that sorbent particle attrition is not detrimental to the combustor performance, but results in improved SO_2 control performance.

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8. PARTICULATE EMISSION CONTROL

Understanding the sources and control of particulate emission from the fluidized-bed combustor is critical to the development of equipment and process designs effective in satisfying the ACL for industrial FBC boilers. In contrast to the ability of wet-scrubber FGD systems to contribute to the control of particulate emissions from conventional coal-fired industrial boiler plants, the injection of sorbent into the fluid-bed combustor for SO_2 capture increases the control requirements of the particulate control equipment.

In this section the effect of boiler design and operating conditions, and of SO_2 emissions control on particulate emissions from industrial FBC boilers, is briefly discussed. The particulate control requirements for a final control device, such as a fabric filter, are estimated. Detailed projections or plant data are not presented, but, rather, a general perspective through illustration of various parameters is developed.

EFFECT OF THE INDUSTRIAL FBC BOILER DESIGN AND OPERATING CONDITIONS

Two configurations of the particulate control system are shown in Figure 15. In the first the primary cyclone has been designed as an integral part of the FBC boiler and functions to collect and reinject relatively coarse carry-over material into the combustor bed for improved carbon and sorbent utilization. The final particulate emission control train consists of a mechanical collector (secondary cyclone) and a fabric filter (baghouse) in series, and would be similar in design for FBC boilers operating with and without primary cyclone recycle (the second configuration shown) of carry-over material to the combustor bed.

The dust loading in the combustion products at the inlet to the final particulate emission control train will depend on FBC boiler design and operating conditions, the coal ash content, the coal and sorbent feed size, and the sorbent particle attrition behavior.

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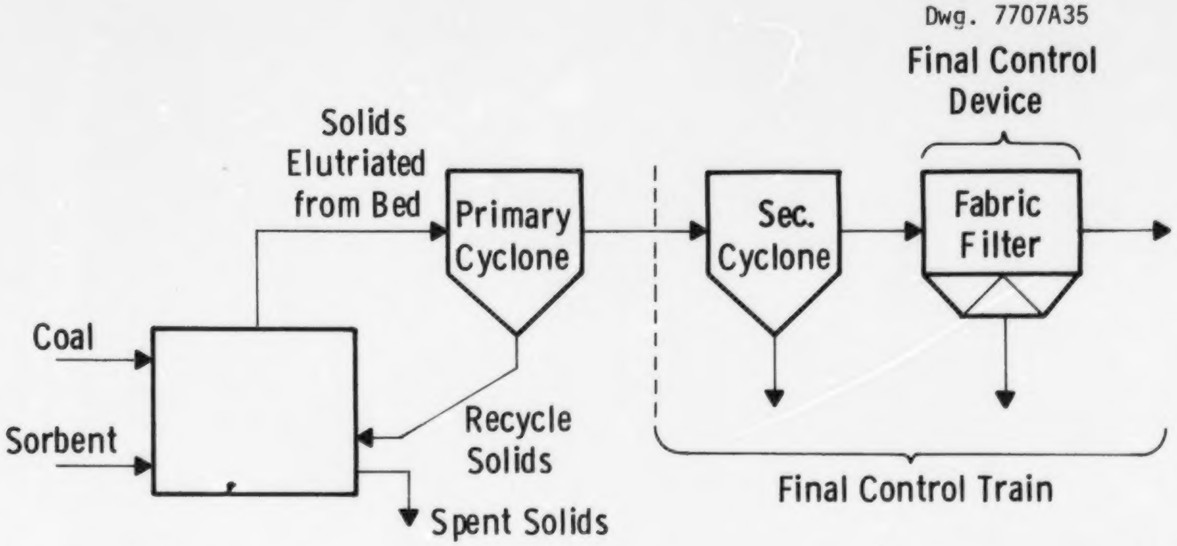
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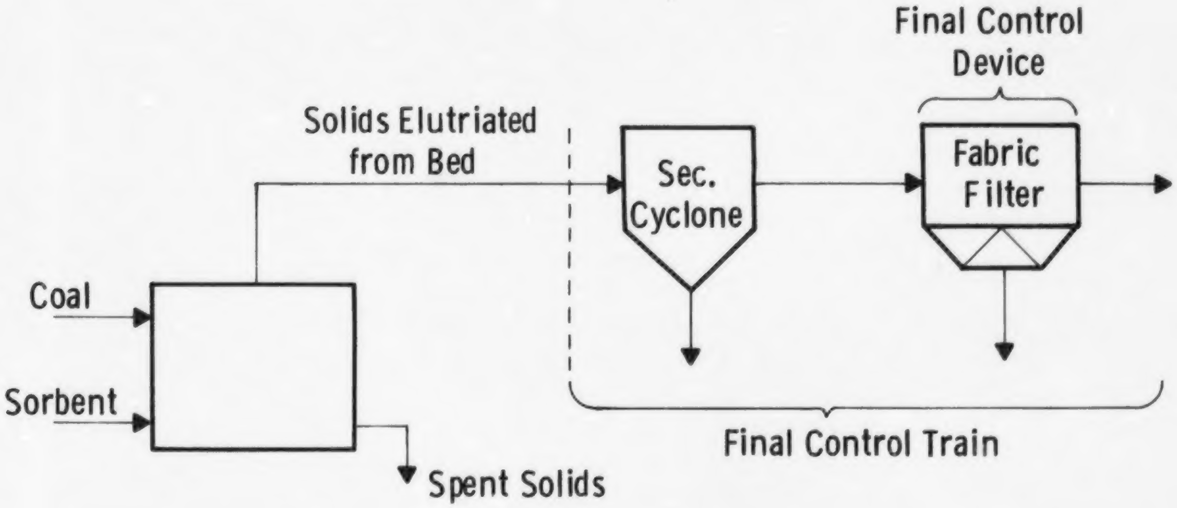
Final Control Device
Solids Elutriated from Bed
Primary Cyclone
Sec. Cyclone
Fabric Filter
Coal
Sorbent
Recycle Solids
Spent Solids
Final Control Train

15a) — Particulate Control System for Industrial FBC Boiler with Primary Cyclone Recycle


Final Control Device
Solids Elutriated from Bed
Sec. Cyclone
Fabric Filter
Coal
Sorbent
Spent Solids
Final Control Train

15b) — Particulate Control System for Industrial FBC Boiler without Primary Cyclone Recycle

Figure 15 - Industrial FBC Boiler Particulate Control

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The solids loading in the gas leaving the fluid-bed combustor will, in general, consist of unburned char, coal ash, and sorbent product particles. The elutriated sorbent material will consist of fine particles originally present in the sorbent feed steam and/or generated by particle attrition processes in the system.

fine

Using a model previously developed by Westinghouse, we have projected size distributions of coal ash, char, sorbent products, and total solids elutriated from utility-size FBC boilers.[5] These projections are shown in Figures 16 through 19. Size distributions very similar to these would be likely from industrial FBC boilers.

In the freeboard, at some distance above the bed surface (the transport disengaging height), the gas velocity tends to become uniform, and the entrained particles consist only of those having terminal settling velocities less than the superficial gas velocity. The shipping constraints (see Table 4) on the design of a shop-fabricated industrial FBC boiler could limit the provision of freeboard height adequate for reaching the transport disengaging height and, hence, could influence the rate and size distribution of particulate emission. The base industrial FBC boiler design places baffles near the bed surface to reduce the necessary freeboard height.

The amount of solids elutriated from the fluid bed is a function of the concentration of fine particles (i.e., particles small enough to be elutriated) in the bed and the fluid-bed operating conditions (such as gas velocity). Fine particles accumulate in the bed through attrition of coarse particles and the feeding or reinjection of fine particles. An increase in any of these variables will increase the solids loading in the flue gas leaving the bed. Table 12 indicates the effect of the FBC boiler design option, recycle versus nonrecycle, on the rate of coal ash and char entering the final particulate control train when 100 percent elutriation of coal ash is assumed. In the case of operation with primary cyclone recycle, improved carbon utilization can result in significantly lower char particle emission. As shown in Figure 20, however, the

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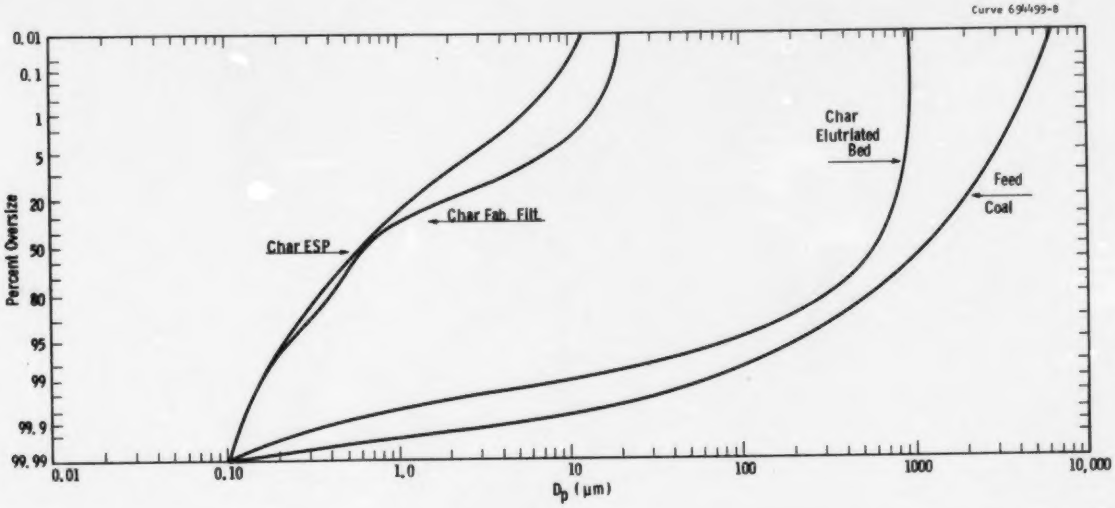

Figure 16 - Coal and Char Size Distributions

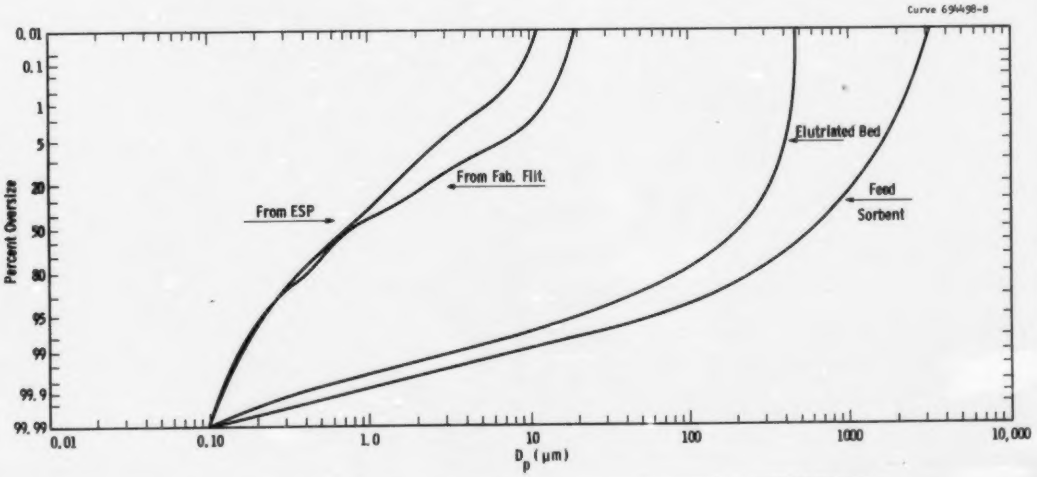

Figure 17 - Sorbent Size Distribution

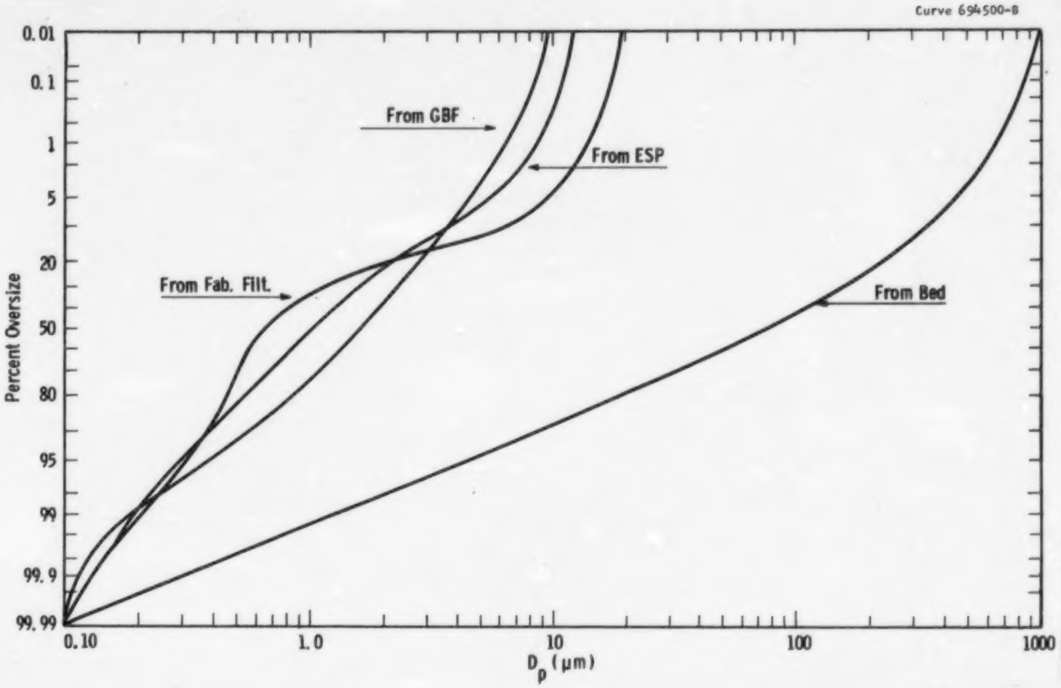

Figure 18 - Size of Ash Elutriated from Bed

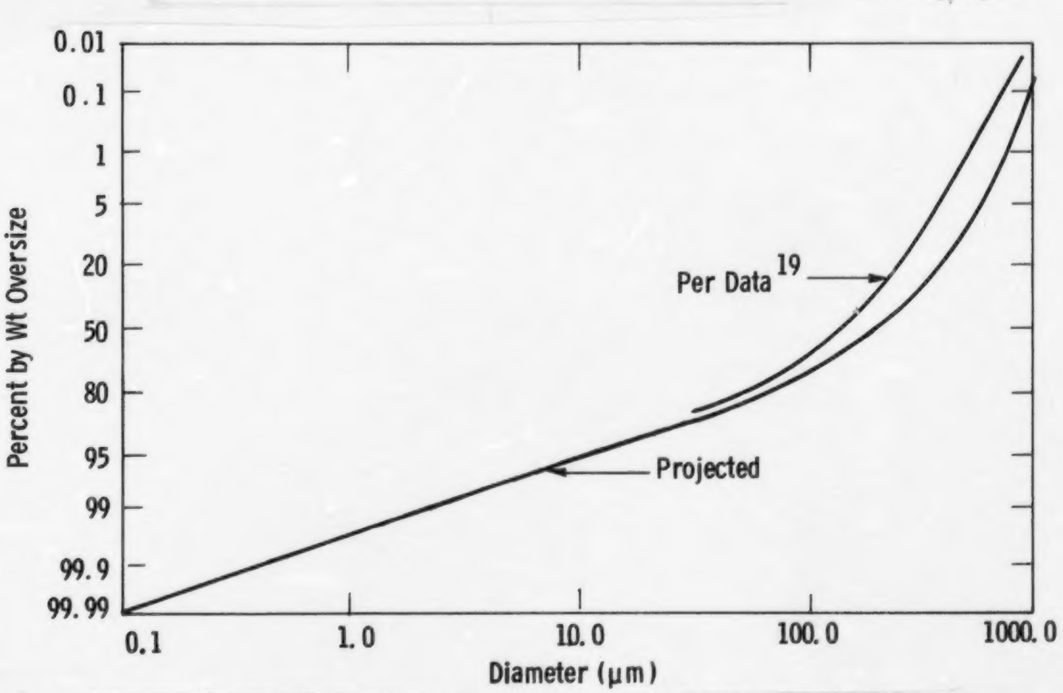

Figure 19 - Total Solids from Combustor

Table 12

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RATE OF COAL ASH AND CHAR ENTERING THE FINAL PARTICULATE CONTROL TRAIN, ng/J (lb/10^6 Btu)

Coal Type		Bituminous High-S	Bituminous Low-S	Western Subbituminous
		Primary Cyclone Recycle Case (95% combustion efficiency)		
Char		1530 (3.6)	1530 (3.6)	1530 (3.6)
Coal Ash		3860 (9.0)	2150 (5.0)	2420 (5.6)
	Total	5390 (12.6)	3680 (8.6)	3950 (9.2)
		Nonrecycle Case (85% combustion efficiency)		
Char		4580 (10.6)	4580 (10.6)	4580 (10.6)
Coal Ash		3860 (9.0)	2150 5.0)	2420 (5.6)
	Total	8440 (19.6)	6730 (15.6)	7000 (16.2)

solids loading in the flue gas passing through the convective pass increases considerably with recycle. The solids loading in the primary cyclone recycle case is 10 to 20 times higher than the nonrecycle case. This increase could, in turn, increase the erosion rate of the heat transfer surface in the convective pass and could influence the maintenance criteria of the fluidized-bed boiler. The dust loading is projected to be twice as great for the high-sulfur coal as it is for the low-sulfur coal.

A typical plot of particle penetration versus particle size for the various particulate densities in a primary cyclone is shown in Figure 21. These data are based on manufacturers' projected performances at temperature.[5]

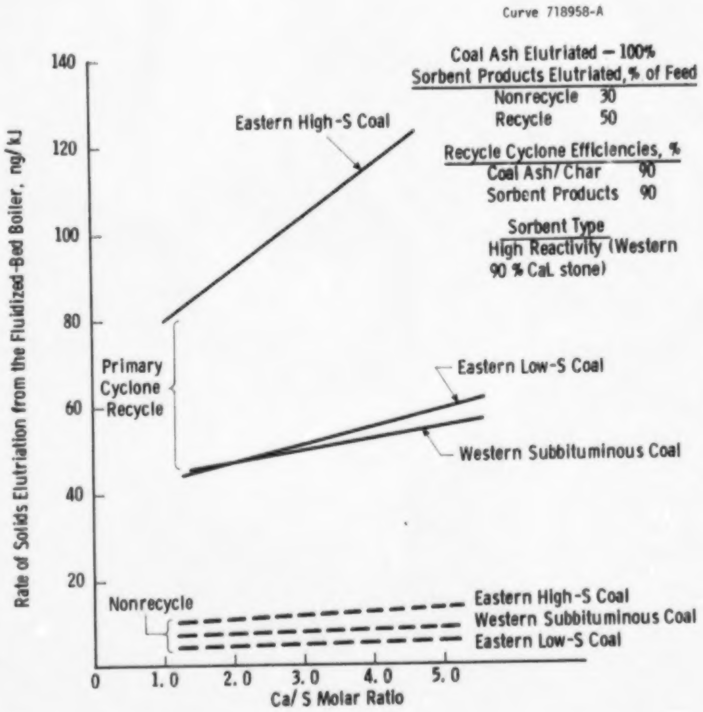

Figure 20 - Rate of Solids Elutriation from the Fluidized-Bed Boiler

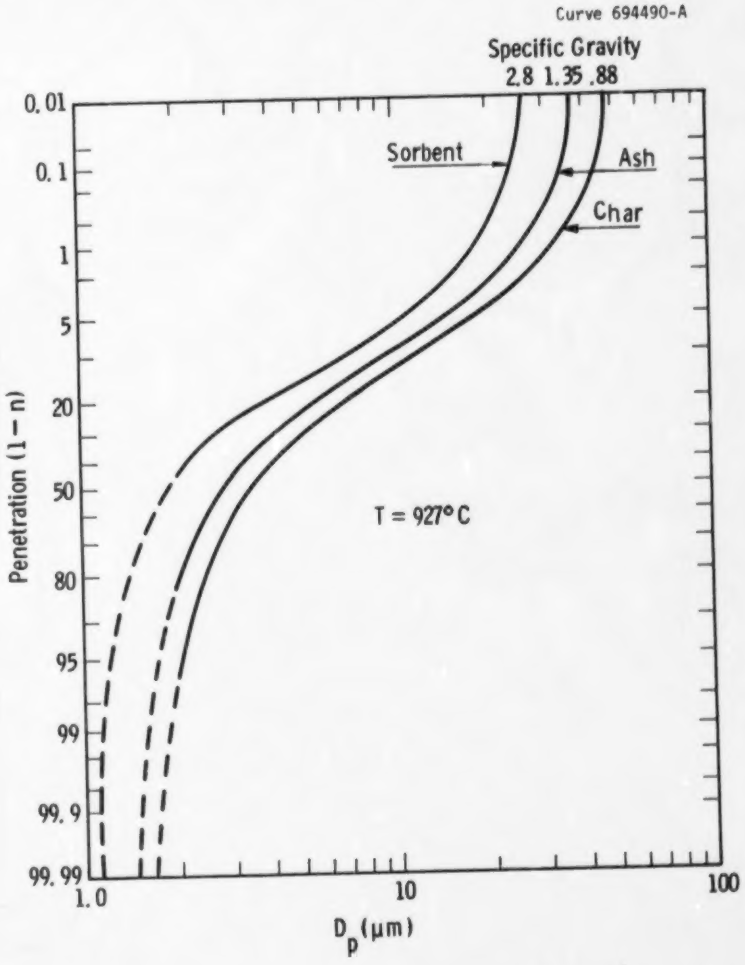

Figure 21 - Grade Efficiency for Primary Cyclone

FINAL PARTICULATE CONTROL TRAIN PERFORMANCE REQUIREMENTS

The particulate collection efficiencies required for the final particulate control device to meet the three alternative particulate control levels are plotted in Figures 22 and 23. We have estimated these efficiencies by assuming the secondary cyclone efficiency to be 80 to 90 percent.

The results plotted in Figure 22 for high-sulfur bituminous coal suggest final particulate device collection efficiency requirements to be in the range of from 85 to more than 99 percent. For the low-sulfur bituminous coal, Figure 23 indicates that a range of from 75 to 99 percent collection efficiency is required. In both figures, regions representing each sorbent activity class for the SO_2 control requirement ranging from moderate to stringent are shown. For the subbituminous coal a range of 74 to 98 percent is projected. The secondary cyclone efficiency and the NSPS level for particulate control have greater influence on the final particulate control requirements than does the Ca/S ratio (sorbent activity and SO_2 control standard).

The maximum mass mean particle sizes entering the final particulate control device could be of the order of 5 to 20 μm, depending on the design of the secondary cyclones. The secondary cyclone is generally designed for inlet gas velocities in the range of 15 to 24 m/s (50-80 ft/sec) and can remove the majority of dust particles above 20 μm in size without excessive loading and wear on the final control device. Figure 24 shows the projected size distribution of the combined solids leaving the primary cyclone for a typical case.

Two major choices for the final stage of particulate removal appear to exist for the industrial FBC boiler: fabric filtration and electrostatic precipitation (ESP), and there is some degree of uncertainty in the use of either of these devices. The precise nature of the electrical properties of the ash, sorbent, and char from the FBC processes is not well known, so migration velocities and possible reentrainment problems in ESP can only be estimated. In view of the considerable problems experienced in trying to precipitate low-sulfur western coals in

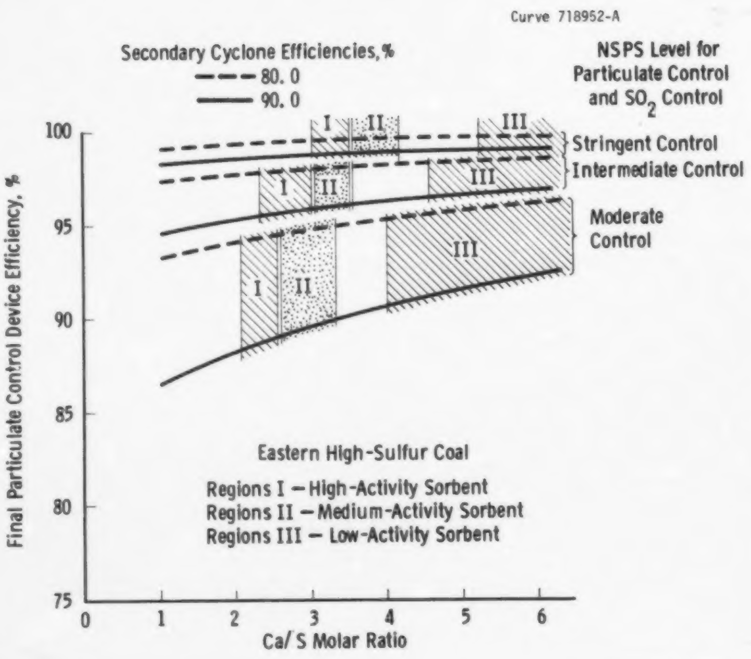

Figure 22 - Final Particulate Emission Control Device Design Requirements for Bituminous High-Sulfur Coal

" Eastern "

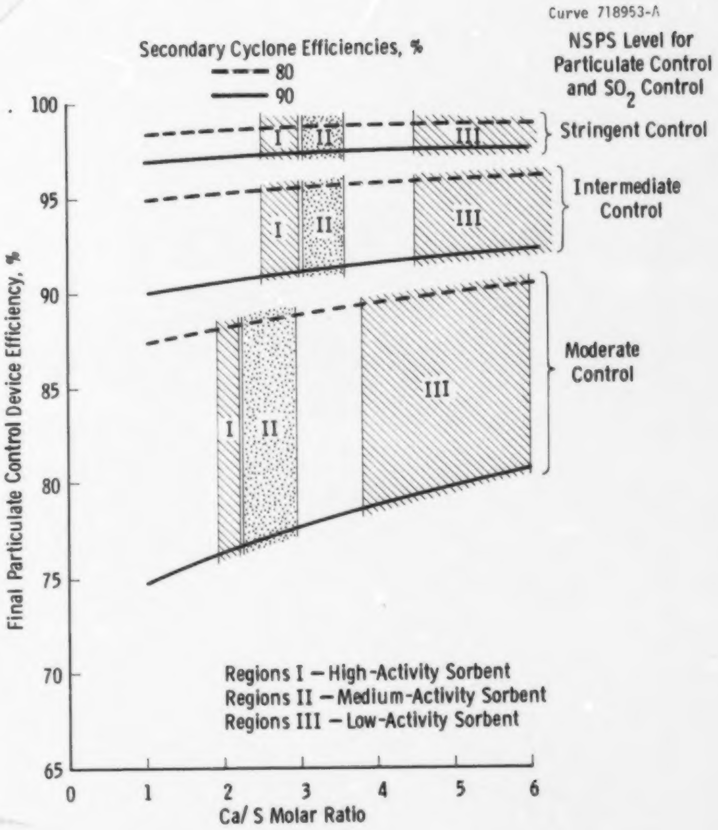

Figure 23 - Particulate Emission Control Devices Design Requirements for Bituminous Low-Sulfur Coal

Eastern "

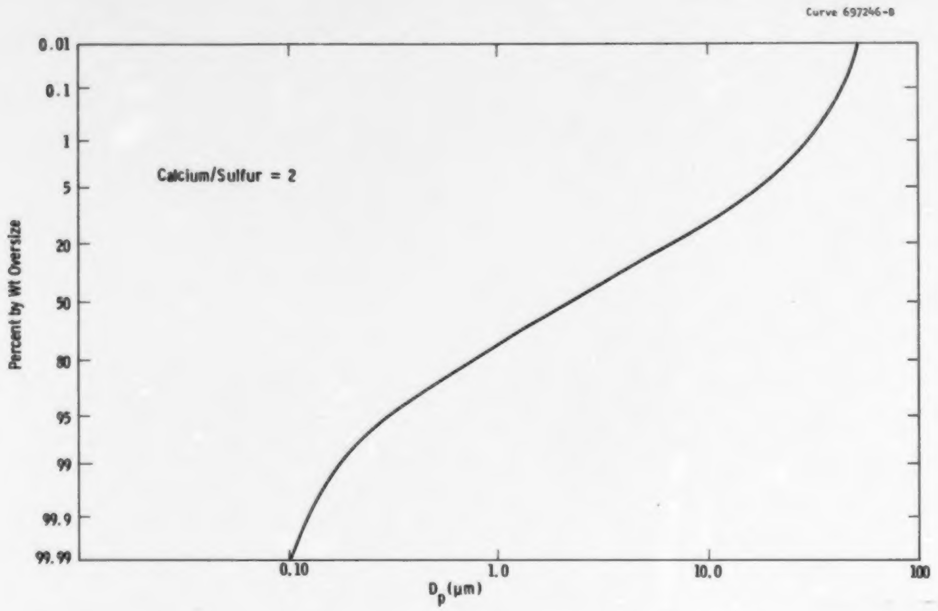

Figure 24 - Combined Solids Leaving Primary Cyclone

conventional boilers, these considerations may be of some consequence in the fluidized-bed case. Grade efficiencies assumed from data reported by J. D. McCain are shown in Figure 25.[23] These data were obtained from measurements on a cold-side pilot-scale ESP operating on a coal-fired boiler burning low-sulfur coal.

The other alternative for the industrial FBC boiler is the use of a fabric filter system. There may be some uncertainty associated with the errosiveness of the particulate and, hence, with bag life, but it seems likely that operating conditions can be chosen to avoid serious maintenance problems with a baghouse system. There is probably less uncertainty about the performance of the filter system than there is with the ESP, since the physical nature of the particles is unlikely to have much effect on the capture efficiency of the filter. Baghouses, however, do have potential problems with start-up and blinding that necessitate careful operation. Grade efficiency has been taken from data reported by H. Spagnola[24] and is shown in Figure 26. These data were collected from a large baghouse facility operating on power plant flue gases.

Both of these final particulate control alternatives may also be subject to fire problems caused by the accumulation of fine carbon particles.

CONCLUSIONS

The particulate control efficiency requirements for the final-stage control device are nearly independent of the industrial FBC boiler plant configuration (primary cyclone recycle versus nonrecycle), the sorbent sulfation activity, and the SO_2 control requirement.

The required efficiency of the final control device is sensitive to the coal properties (sulfur content and ash content) and the efficiency of the secondary cyclone.

As with sulfur removal control, the combustor design and operating conditions greatly affect the nature of particulate emissions. The combustor can be designed and operated (e.g., at low velocities) to limit particle attrition and elutriation, though rather efficient final-stage particulate control devices will still be required (87 to 99.5 percent for high-sulfur bituminous coal; 74 to 98 percent for the low-sulfur subbituminous coal).

The major uncertainty associated with the primary cyclone recycle option is the high dust loadings expected in the boiler convective zone and the resulting possibility of high erosion rates on the convective heat transfer surface. Placement of the convective surface following the primary cyclone may be required.

Conventional technology exists - ESP or fabric filter techniques (baghouses) - that should be capable of controlling particulate emissions to the most stringent level of 12.9 ng/J (0.03 lb/MBtu). Testing these control options on operating fluidized-bed combustors, however, is

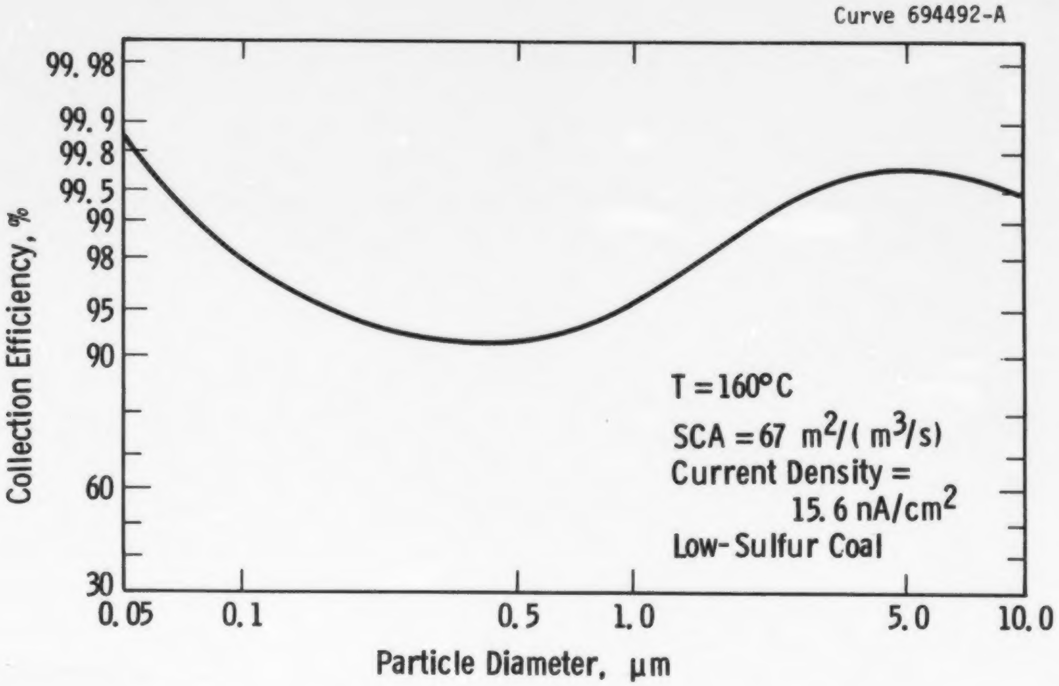

Figure 25 - Grade Efficiency for ESP (Ref 17)

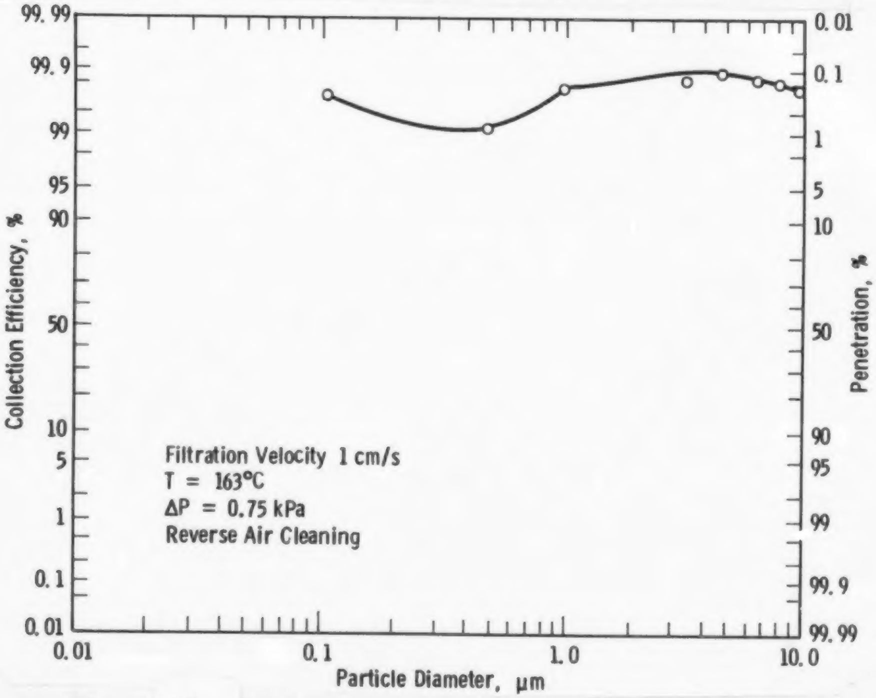

Figure 26 - Baghouse Performance at Sunbury Steam Electric Station (Ref 18)

Curve 694489-A

necessary before actual control performance and any operating problems can be firmly defined. No such testing has yet been conducted.

Fabric filter techniques seem appropriate for the industrial FBC boiler, with the major uncertainties involving the details of the equipment specifications (filter materials, face velocities, etc.), operating and maintenance procedures, and potential problems with carbon accumulation and baghouse fires.

The nature of submicron particles emitted from the final particulate control device may become an important issue in the future when such emissions may be regulated and may require special control equipment. Present data, and the scope of this evaluation, do not permit a comparison of the potential environmental concerns with regard to submicron particle emissions from AFBC and conventional boilers.

9. SOLID WASTE GENERATION

In this section the impact of the alternative potential levels of SO_2 control in industrial FBC boilers is discussed with respect to solid waste generation. Both the quantity of material produced and its nature are considered, and the processing and disposal of solid waste is assessed.

The disposal of solid waste is governed by laws promulgated under the 1976 Resource Conservation and Recovery Act (RCRA, PL 94-580). The EPA plans to propose regulations governing special wastes in the early part of 1982. FBC solid residue, by virtue of the large quantities produced, seems to be a candidate for inclusion in the special wastes category, which will include solid residues from cement kilns, utility power plants, phosphate mining and processing, uranium mining and processing, oil and gas drilling muds, and the solids from other mining operations.

The waste solids produced by industrial FBC boiler systems consist of combustor bed overflow material and the elutriated particulate material collected by the particulate control devices. The nature of the waste solids depends on the coal and sorbent characteristics, plant operating conditions, and SO_2 emissions control standard being achieved. Particulate matter collected by the particulate control devices can be mixed with the bed overflow material or can be handled separately. In general, the spent bed material will consist of partially sulfated sorbent particles and small amounts of accumulated coal ash, whereas the elutriated fines will be rich in coal ash material and will have a substantial content of partially sulfated and partially calcined sorbent particles, and unburned carbon. The chemical composition, size distribution, and physical properties of these two waste solids streams can be significantly different.[25] Sorbent attrition and feed size distribution characteristics significantly affect the distribution of waste sorbent particles between the bed overflow material and the elutriated fines. For a given plant size the total amount of spent solids produced will depend on the coal and sorbent feed rates and their chemical composition.

The material handling requirements for the elutriated particulate can be significantly different from that for the relatively coarse and hotter bed overflow material. Fluidized-bed operating conditions like superficial velocity, bed depth, and sorbent feed particle size distribution can have a significant effect on the amount of solids elutriated and, hence, can indirectly influence the waste solids handling system requirements.

BASE DESIGN WASTE SOLIDS

The total amount of solid residue produced relative to the coal feed rate (consisting of the bed overflow material and the elutriated fines) for the three reference coals at the three SO_2 emission control levels, using 500 and 1000 μm average bed sorbent particle sizes, are summarized in Tables 13 and 14, respectively.

In the case of low-sulfur coals, the waste solids consist primarily of unburned carbon and coal ash, and the total amounts produced are 8 to 26 percent of the coal feed rate, comparable to that from conventional coal-fired boilers. The increased sorbent requirements for the bituminous high-sulfur coal, however, result in approximately 36 to 81 percent solid waste, depending on the sorbent reactivity and SO_2 emission control level, and this is significantly more than the amount of (50 percent moisture) sludge produced from a conventional boiler plant with a lime wet-scrubbing system.

FEASIBILITY OF PROCESSING INDUSTRIAL FBC BOILER SOLID WASTE

The solid residue produced by the industrial FBC systems could be processed for utilization and/or to minimize the environmental impact of disposal. Westinghouse has considered several processing options, in addition to evaluating the environmental impact of direct disposal.[26]

Table 13

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TOTAL WASTE SOLIDS WITH 500 μm AVERAGE BED PARTICLE DIAMETER

Sorbent Activity Category	High (Western 90% CaL Stone)	Medium (Bussen Quarry)	Low (Menlo Quarry)
	Mass/Mass of Coal Feed		
SO_2 Emissions Control Standard (% S removal)	Bituminous High-S Coal (3.5 wt.% S, 10.58 wt.% ash)		
Stringent (90)	0.42	0.53	0.59
Intermediate (85)	0.39	0.48	0.54
Moderate (78.7)	0.36	0.44	0.50
	Bituminous Low-S Coal (0.9 wt.% S, 6.9 wt.% ash)		
Stringent & Intermediate (84.7)	0.18	0.20	0.22
Moderate (75)	0.17	0.19	0.20
	Western Subbituminous Coal (0.6 wt.% S, 5.4 wt.% ash)		
Stringent & Intermediate (84)	0.09	0.10	0.11
Moderate (75)	0.08	0.09	0.10

Table 14

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TOTAL WASTE SOLIDS WITH 1000 μm AVERAGE BED PARTICLE DIAMETER

Sorbent Activity Category	High (Western 90% CaL Stone)	Medium (Bussen Quarry)	Low (Menlo Quarry)
	Mass/Mass of Coal Feed		
SO_2 Emissions Control Standard (% S removal)	Bituminous High-S Coal (3.5 wt.% S, 10.58 wt.% ash)		
Stringent (90)	0.46	0.61	0.81
Intermediate (85)	0.42	0.55	0.73
Moderate (78.7)	0.38	0.49	0.66
	Bituminous Low-S Coal (0.9 wt.% S, 6.9 wt.% ash)		
Stringent & Intermediate (84.7)	0.19	0.22	0.26
Moderate (75)	0.17	0.20	0.24
	Western Subbituminous Coal (0.6 wt.% S, 5.4 wt.% ash)		
Stringent & Intermediate (84)	0.09	0.10	0.11
Moderate (75)	0.08	0.09	0.10

CAUTION: Mark corrections in Sky Blue pencil ONLY

The objectives of the various processing options are to reduce the environmental activity of the waste solids by:

- Physical changes, as in dead-burning
- Chemical changes, as in wet carbonation
- Combinations of physical and chemical changes, as in ash blending.

The possible end uses considered for the processed material include:

- Soil stabilization
- Agricultural uses
- Landfill
- Concrete/aggregate
- Gypsum
- Mineral recovery
- Refractory brick
- Acid mine drainage treatment
- Municipal waste treatment.

On the basis of exploratory work, Westinghouse has concluded that ash blending is technologically the most promising FBC solid residue processing option.[27] Preliminary experimental tests have revealed that stable solid compacts can be produced by grinding FBC spent sorbent to <125 μm (-120 mesh), combining it with FBC carry-over material, and adding water. These solid compacts seem to be similar to low-strength concrete and high-strength aggregate. Thus, the spent solids can be cast in the form of monolithic blocks for disposal (landfill) in an environmentally acceptable manner and/or produced in any other shape suitable for use in the construction industry. The economic feasibility of processing waste solids, especially from relatively small plants like the industrial-sized units considered in this study, is yet to be demonstrated and is questionable.

WASTE SOLIDS DISPOSAL OPTIONS

Disposal on land or in the ocean are two disposal alternatives available. In the case of land disposal several options exist:

- Below grade (e.g., worked-out mines and quarries)
- At grade (e.g., valleys, artificial ponds)
- Above grade (e.g., stockpiles).

The applicability of a specific disposal option will depend on several factors, such as availability of the disposal site to the FBC plant, population centers, groundwater facilities, and the potential environmental impact.

ENVIRONMENTAL IMPACT

Westinghouse[5] has performed laboratory-scale investigations on solid residue from FBC test facilities PER, Argonne National Laboratories [ANL], Exxon, National Coal Board [NCB], etc.) and has concluded that the waste solids will not be a hazardous pollutant. Because of the large amounts to be handled, however, the residue is still a candidate for the RCRA special waste category. Leaching and activity tests for spent solids from limestone-based sorbents revealed that:

- FBC residues tested would not constitute a hazardous waste on the basis of RCRA criteria.
- The leachate meets Drinking Water Standards (DWS), and, thus, no drinking water pollution problem exists.
- The sulfide concentration and the total dissolved organics (TOC) are below detection limits.
- The large amount of CaO present in the spent solids is of environmental concern because of the heat release properties.
- The high concentrations of calcium, sulfate, total dissolved solids (TDS), and the pH, which are above the DWS are of potential concern.

- The addition of $\geq$20 wt% coal ash to the bed overflow waste seems to improve leachate quality, and, therefore, codisposal of bed overflow material and elutriated fines may be desirable.
- Room-temperature processing, such as bed overflow material and elutriated particulate blending in the presence of water, followed by air curing to produce solid compacts, seems to improve leachate quality and minimize environmental impact.

COMPARISON WITH FGD RESIDUES

A comparison of the leachate characteristics of FBC and FGD (lime/limestone wet-scrubbing) residues is presented in Table 15. Those species that exceed the DWS are cross-hatched to put leachate quality in perspective. The presence of calcium sulfite ($CaSO_3$), relatively higher concentrations of magnesium, chlorine, arsenic, selenium, cadmium, manganese, and fluorine, and the difficulties in transporting and handling the FGD sludge from conventional boilers suggest more negative environmental effects than from the disposal of dry, granular, waste residue from FBC boiler systems. Table 16 summarizes the results from the comprehensive test program. The results with natural gypsum are also shown for perspective.

CONCLUSIONS

The variation in the three levels of control for SO_2 has an insignificant influence on the mass rate of waste solids produced by industrial FBC boilers at the base design conditions (<23% change). The sorbent activity has greater impact on the mass rate of waste solids production for the high-sulfur eastern coals (up to 77% change). For the low-sulfur coals the ACL has an even less significant influence. Selection of proper operating and design conditions for the industrial FBC boiler can reduce the solid waste production rate, the lower activity sorbent being more sensitive to those factors.

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Table 15

COMPARISON OF LEACHATE CHARACTERISTICS FROM THE FBC AND FGD RESIDUES

Substance	Liquor, mg/ℓ	Leachate, mg/ℓ		Drinking Water* Standards, mg/ℓ
	FGD	FBC	FGD	
Al	0 to 20	0 to > 2	< 1	
Ag	< 0.05	< 0.05	< 0.05	0.05
As	< 0.05	< 0.05	▨ 0 to 0.1	0.05
B	> 5	0 to > 5	> 1	
Ba	< 1	▨ < 2**	< 1	1.0
Be	< 0.02	< 0.02	< 0.02	
Bi	< 0.04	< 0.04	< 0.04	
Ca	▨ > 500	▨ > 500	▨ > 500	75
Cd	▨ 0 to 0.2	< 0.01	< 0.01	0.01
Co	< 0.1	< 0.1	< 0.1	
Cr	< 0.05	▨ < 1.0**	< 0.05	0.05
Cu	< 1	< 0.1	< 0.1	1.0
Fe	< 0.3	< 0.3	< 0.3	0.3
Hg	< 0.002	< 0.002	< 0.002	0.002
Mg	▨ 0 to > 1000	▨ 0 - 250**	▨ 0 to 500	50
Mn	▨ 0 to 20	< 0.05	▨ 0 to 0.1	0.05
Mo	0.1 to 7.0	< 5	< 1	
Na	0 to > 100	0 to > 100	< 10	
Ni	< 1	< 0.1	< 0.1	2.0
Pb	< 0.05	< 0.05	< 0.05	0.05
Sb	< 0.2	< 0.1	< 0.1	
Se	▨ 0.001 to 0.5	< 0.01	▨ 0 to 0.1	0.01
Si	0 to 30	0 to 30	0 to 5	
Sn	< 1.0	< 1.0	< 1.0	1.0
Sr	0 to 40	0 to > 10	0 to 5	
Ti	< 2	< 2	< 2	
V	< 2	< 1	< 1	
Zn	< 2	< 1	< 1	5.0
Zr	< 2	< 1	< 1	
SO_3	< 10 to 40	< 10	< 10	
SO_4	▨ 1000 to 7000	▨ 1000-2000	▨ 1000-2000	250
Cl	▨ 300 to 6000	▨ 0 to 350**	▨ 30 to 300	250
F	▨ 10 to 50	< 2.4	▨ 1 to 10	2.4
NO_3(as N)	▨ 0 to 100	< 10	< 10	10
TOC	< 30	< 30	< 30	
pH	6 to 9	▨ 9 to 12	6 to 9	5 to 9
TDS	▨ 5000 to 14000	▨ 2000 to 4000	▨ 2000 to 3000	500
Specific Conductance µmho/cm	5.0 to 17.0	0.5 to 10.0	2.0 to 3.0	

* NIPDWR and USPHS (1962) Drinking Water Standards, and WHO Potable Water Standards

** Concentrations Higher Than the Drinking Water Standards Resulted from Leachates of < 2 Batches of Carry-over Fines Among the > 30 Spent FBC Materials Investigated

▨ Exceed Drinking Water Standards

The production of solid waste by the industrial FBC boiler burning high-sulfur eastern coal is from once to twice the amount of sludge (50% water) produced by a conventional boiler with lime/limestone wet-scrubbing. For the low-sulfur coals the waste solid from the industrial FBC boiler is comparable in production rate to the conventional industrial boiler sludge. FBC solid residue disposal by landfill seems to be the most promising option, with relatively less environmental impact than the FGD sludges from conventional coal-fired plants. The waste solids, depending on economic limitations, can be processed to produce marketable material for the construction industry. This strategy seems unlikely for industrial FBC boilers because of their limited capacities relative to market needs.

On the basis of current information, AFBC residue can be considered nonhazardous and, therefore, will be disposed of in a sanitary landfill. Regulations and disposal criteria for sanitary landfills under the RCRA Secs. 4004 and 1008 and the Clean Water Act will determine FBC disposal practice.

Since the groundwater criteria, applicable at the "disposal facility boundary," will incorporate effects of soil attenuation and dilution, these effects should be investigated further. These effects are especially important with respect to leachate pH, TDS, sulfate, and calcium, should the secondary DWS be included as part of the groundwater criteria in the future. Control technologies for major leachate variables, i.e., pH, TDS, sulfate, and calcium, should be evaluated in the event that dilution/attenuation effects might be insufficient for some specific sites.

Other aspects of disposal, such as the proper procedures for handling the FBC residue, disposal site selection, facility design, construction, operation, and management need to be investigated.

Table 16

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COMPARISON OF ENVIRONMENTAL CONCERNS

Process	FBC		FGD[b]			Natural Gypsum
Parameter	Unprocessed	Processed	Unprocessed	Oxidized	Stabilized	Unprocessed
Hazardous Characteristics[a]	no	no	no	no	no	no
Leachate Characteristics						
Trace Elements			X	X		
Total Organic Carbon						
pH	X	c				
TDS	X	c	X	X	X	X
Ca	X	c	X	X	X	X
Mg			X	X	X	
Sulfate	X	c	X	X	X	X
Sulfite			X			
Sulfide	d					
Thermal Activity	yes[e]	no	no	no	no	no
Physical Strength	no	no	yes	no	no	no
Quantity	yes	yes	yes	yes	yes	Not applicable

X = Exceeds DWS where they exist.

[a] Detailed report on test data will be issued later in 1980.

[b] Limited to lime or limestone scrubber sludge.

[c] Leachability significantly reduced, degree of improvement depending on processing alternatives.

[d] Not a concern for the once-through FBC residue; further investigation needed for spent sorbent with sorbent regeneration.

[e] Associated with residue with a substantial amount of unutilized CaO, e.g. AFBC bed material.

Table 16
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10. NITROGEN OXIDE EMISSIONS CONTROL IN INDUSTRIAL FBC BOILERS

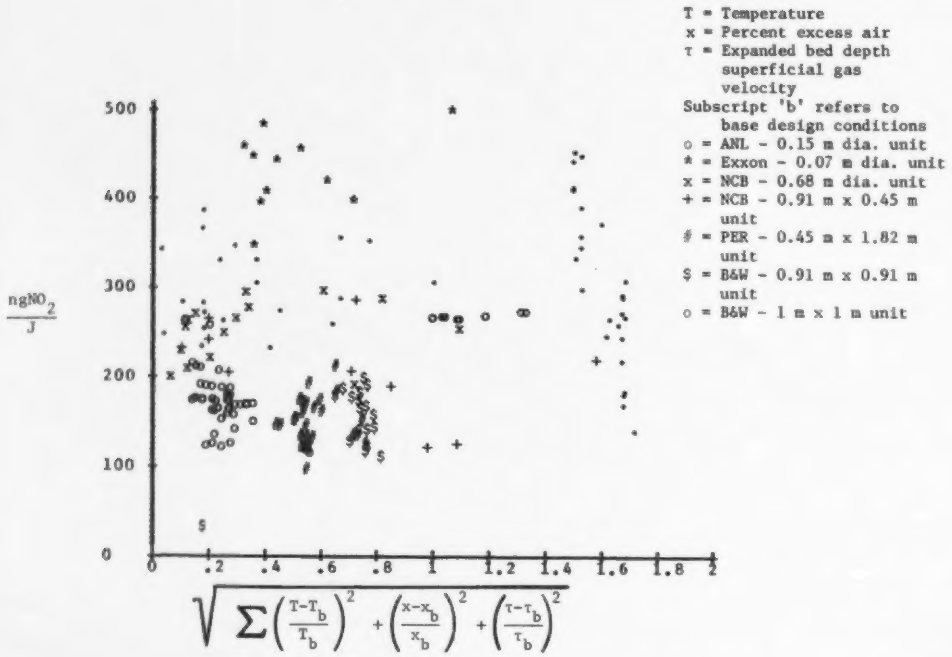
Nitrogen oxide emissions from developmental fluidized-bed combustors have been generally observed to be significantly lower than those from conventional coal-fired boiler systems without special combustion modifications. This difference is due partially to FBC operation at relatively low combustion zone temperatures and with less excess air than many types of conventional industrial boiler units.

For the control of SO_2 and particulates, the control techniques and relation to FBC operating and design parameters are relatively clear. For NO_x, however, the relation between emissions and operating and design parameters in FBC are not clearly understood. Efforts are under way to improve understanding of these relationships and to develop NO_x control strategies that can reduce FBC NO_x emissions below their already inherently low levels.

Data on NO_x emissions from several FBC units are summarized in Figures 27 and 28. These data represent NO_x emissions over a broad range of temperatures, excess air levels, gas residence times, and coal nitrogen contents. The data are presented as scatter plots, as suggested by MIT investigators.[28] Figure 27 shows the NO_x emission as ng NO_2/J of fuel input versus the "distance" from the base design point (840°C, 20% excess air, gas residence time of 0.675 seconds). From this plot, if we exclude the data from the extremely small ANL and Exxon units, and draw an envelope around the data points, the NO_x emission at the base design point would be expected to be between 150 and 250 ng NO_2/J, with a mean of about 200 ng/J.

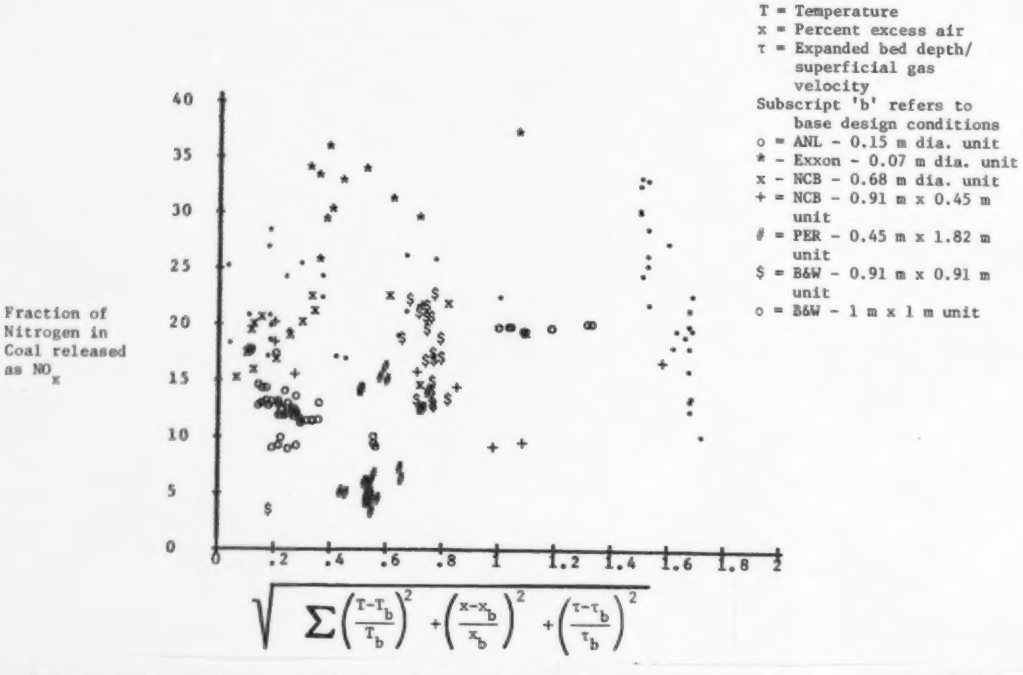
NO_x ^

Figure 28 presents the same data as a fraction of the coal nitrogen released as NO_x. The plot suggests that 12 to 18 percent of the coal



$$\sqrt{\Sigma\left(\frac{T-T_b}{T_b}\right)^2+\left(\frac{x-x_b}{x_b}\right)^2+\left(\frac{\tau-\tau_b}{\tau_b}\right)^2}$$

Figure 27 - Scatter Plot of Total NO_x Emission Data



$$\sqrt{\Sigma\left(\frac{T-T_b}{T_b}\right)^2+\left(\frac{x-x_b}{x_b}\right)^2+\left(\frac{\tau-\tau_b}{\tau_b}\right)^2}$$

Figure 28 - Scatter Plot of NO_x Emission Data as Fraction of Coal Nitrogen

nitrogen will be emitted as NO_x in the base design case. Using this range of expected behavior, Table 17 gives the expected NO_x emissions for the three reference coals at the base design conditions.

Table 17

PROJECTED NO_x EMISSIONS AT THE BASE DESIGN CONDITIONS

	ng NO_2/J		
	Percent Coal Nitrogen Released as NO_x		
Coal	12	15	18
High-Sulfur Eastern	187	233	280
Low-Sulfur Eastern	184	230	277
Western Subbituminous	212	265	318

link table heads only

Table 17 indicates that there should be little difficulty in satisfying the moderate NO_x control level of 310 ng/J at the base conditions with the reference coals. Several options are available to improve the NO_x emission, ranging from simply reducing the excess air level to incorporating staged combustion into the unit design. While the moderate and intermediate NO_x control levels should be inherently achievable with many combustor designs and many coals or by suitable modifications to the basic combustor operating conditions, the stringent control level of 215 ng/J, may well require in many cases a more sophisticated control technique. Various control techniques are under investigation, but their performance and impact on the total plant performance are not yet known.[29,30]

Statistical correlations have also been developed for industrial-scale AFBC NO_x emissions.[31] Such a correlation, based on most of the available PFBC and AFBC data, is:

$$\begin{aligned} NO_x = &- (0.5775)(P) + (4.135 \times 10^{-4})(P)^2 + (0.6598)(T) \\ &- (4.238 \times 10^{-4})(T)^2 + (1.1668)(X) \\ &- (2.759 \times 10^{-3})(X)^2 - (5.5066)(h/v) \ , \end{aligned}$$

where NO_x is expressed as ng NO_2/J, and

P = pressure, kPa

T = temperature, °C

X = percent excess air

h = bed depth, m

v = superficial gas velocity, m/s.

The scatter around this correlation is rather large, but it provides an alternative to the scatter plot approach for estimating whether the selected operating conditions will result in acceptable NO_x emissions and a basis for selecting modified conditions.

Modeling approaches to the prediction and correlation of NO_x emission behavior are currently being developed.

Pereira and Beér[32] have developed a model describing the formation and destruction of NO in homogeneous reactions involving the coal volatiles. The model consists of two parts - the homogeneous reaction process of NO formation and destruction and the heterogeneous process of NO formation and the NO-char reduction. Simulated results from the theoretical model were observed to agree reasonably well with experimental data for temperatures between 600 and 800°C. Above 800°C, however, the discrepancies between model predictions and the experimental results were attributed to the lack of information on the fast reduction of NO by char.

Most of the researchers investigating NO_x emissions in FBC have reported significantly higher concentrations of NO near the distributor plate (air inlet to the bed) and a decrease in NO concentration as the gas rises through the bed. NO-char reduction reactions at temperatures of 750°C and higher have been suggested to be primarily responsible for the decomposition of the once-formed NO. In order to provide a preliminary test of this mechanism, we calculated kinetic rate constants for NO_x decomposition from available AFBC data by assuming that all the nitrogen in the coal is released at the base of the bed and that the NO_x then decomposes according to first-order kinetics as it passes up through the depth of the bed (to reach the concentrations measured in the flue gas). The scatter in the first-order specific reaction rate constants plotted in Figure 29 shows the inadequacy of the first-order decomposition model. These specific reaction rate constants, however, are within the range reported by NO-char reduction studies at MIT.

Thus, the modeling approach has not quite reached the level of development where confident projections can be made that relate the design and operating conditions to NO_x emission performance. Trends in performance, however, can be properly estimated.

In contrast with the sulfur removal case, in which the emission standard is an important factor but not critical to the feasibility of AFBC dense-bed boilers, the NO_x emission standard could be critical to the successful commercialization of industrial AFBC dense-bed boilers. If stringent standards are imposed, requiring lower bed temperatures, lower excess air levels, coarser coal feed sizes, or two-stage combustion for compliance, then the impact on the corrosion of heat transfer surfaces and the combustion efficiency of the boiler may become limiting considerations.

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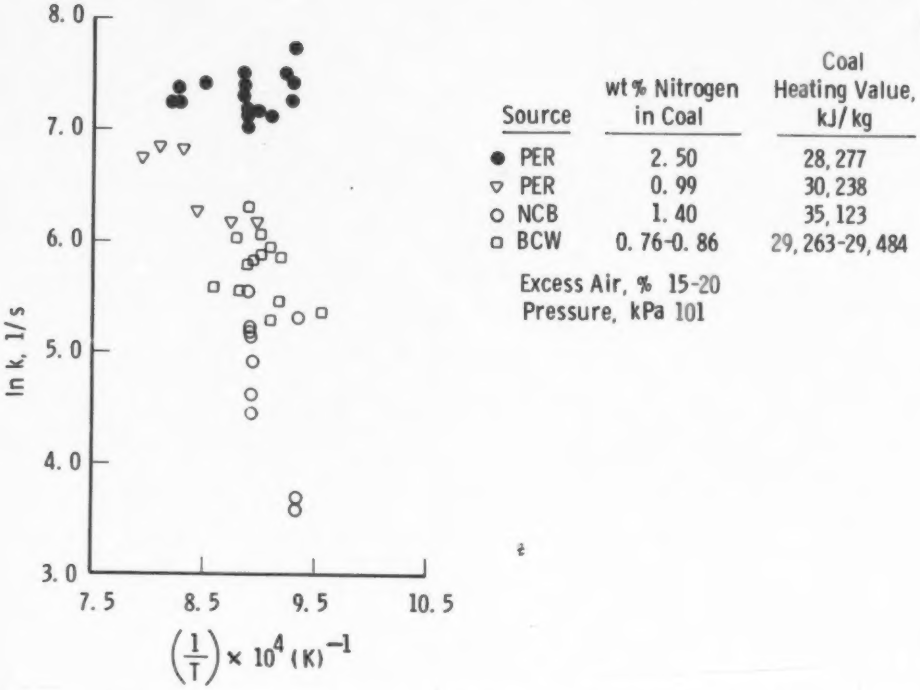

Figure 29 - NO Decomposition Model - Pseudo-First-Order Rate Constants

11. PLANT ENERGY CONVERSION EFFICIENCY

This section discusses the effect of the ACL on industrial FBC boiler energy conversion efficiency and includes:

- An assessment of the base industrial FBC boiler design performance
- A parametric evaluation of the effects of plant design and operating conditions
- A comparison with conventional coal-fired industrial boiler plants using FGD devices for SO_2 control.

In this study boiler and plant efficiencies have been defined on the basis of the thermal input to the boiler. The boiler efficiency accounts for inherent energy losses in the boiler (steam energy divided by fuel input energy), whereas the plant efficiency accounts for all forms of energy input to the plant (steam energy divided by total plant energy input).

BASE DESIGN PERFORMANCE

The impact of the ACL on industrial FBC boiler energy conversion efficiency will be chiefly through the energy losses due to sorbent calcination, escape of unburned combustibles, and the unrecovered sensible and latent heat of the flue gas components to the stack. These losses are summarized in Tables 18 and 19 for the cases of 500 and 1000 μm average bed particle sizes, respectively, and are discussed below.

Combustion Efficiency

Combustion efficiency accounts for the energy losses due to the escape of unburned combustibles from the boiler, primarily in the form of char, with considerably smaller contributions from combustible gases, primarily carbon monoxide (CO), and lighter hydrocarbons. The FBC combustion

Table 18

Dwg. 7707A32

BASE INDUSTRIAL FLUIDIZED-BED BOILER ENERGY CONVERSION PERFORMANCE: 500 μm AVERAGE BED PARTICLE DIAMETER

Losses, % of thermal input	Coal Type		
	Bituminous High-S	Bituminous Low-S	Western Subbituminous
Unburned Carbon	5.00	5.00	5.00
Dry Flue Gas, 135°C (275°F) Stack Inlet	4.40	4.40	4.42
Water and Hydrogen in Coal	4.71	3.75	5.89
Radiation and Unaccounted	1.50	1.50	1.50
Calcination and Sulfation*	0.09 - 2.18	0.01 - 0.39	−0.16 - 0.07
Total Losses	15.7 -17.79	14.66-15.04	16.65 - 16.88
Boiler Efficiency, %	84.30-82.21	85.34-84.96	83.35 - 83.12
Auxiliary Power Consumption,** % of thermal input	3.39- 3.92	3.11 - 3.21	3.06 - 3.12
Plant Efficiency, %	80.91 -78.29	82.23 -81.75	80.29 - 80.00

* Range for the various sorbents and SO_2 emission control levels; lower limit- high-activity sorbent (Western 90% CaL stone) @ moderate control; upper limit - low-activity sorbent (Menlo quarry stone) @ stringent control. Corresponds to Ca/S = 2.1 - 5.3 respectively, for high-sulfur bituminous coal.

**Assumes electricity is supplied by a power station with 33 % efficiency.

Table 19

Dwg. 7707A33

BASE INDUSTRIAL FLUIDIZED-BED BOILER ENERGY CONVERSION PERFORMANCE 1000 μm AVERAGE BED PARTICLE DIAMETER

Losses, % of thermal input	Coal Type		
	Bituminous High-S	Bituminous Low-S	Western Subbituminous
Unburned Carbon	5.00	5.00	5.00
Dry Flue Gas, 135°C (275°F) stack inlet	4.40	4.40	4.42
Water and Hydrogen in Coal	4.71	3.75	5.89
Radiation and Unaccounted	1.50	1.50	1.50
Calcination and Sulfation*	0.37 - 4.55	0.06 - 0.82	−0.13 - 0.17
Total Losses	15.98-20.16	14.71-15.47	16.68 - 16.98
Boiler Efficiency, %	84.02-79.84	85.29-84.53	83.32- 83.02
Auxiliary Power Consumption,** % of thermal input	3.15 - 3.39	3.05 - 3.11	3.03 - 3.06
Plant Efficiency, %	80.87 -76.45	82.24 -81.42	80.29 -79.96

* Range for the various sorbents and SO_2 emission control levels; lower limit - high-activity sorbent (Western 90% CaL stone) @ moderate control; upper limit - low- activity sorbent (Menlo quarry stone) @ stringent control. Corresponds to Ca/S = 2.5 - 8.5, respectively for high-sulfur bituminous coal.

**Assumes electricity is supplied by a power station with 33 % efficiency.

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efficiency is dependent on the coal properties, the coal feed size distribution, the coal feed system design (location and distribution of coal feed points), the bed operating conditions (temperature, superficial velocity, excess air, bed depth), and the freeboard design (above-bed combustion of volatiles and CO occurring in the region below the convective cooling region). Combustion efficiencies of the order of 95 to 97 percent and 85 to 96 percent have been reported for AFBC units operating with and without recycle of carry-over fines, respectively.[20,25,33,34] Significantly higher combustion efficiencies would be expected for the subbituminous coals than for the bituminous coals. In this study the boiler efficiency has been calculated by assuming 95 percent combustion efficiency with reinjection of carry-over fines into the combustor bed for improved carbon utilization. In conventional spreader stokers with recycle and in pulverized-coal-fired units, combustion efficiencies of 95 to 97 percent, and more than 99 percent, respectively, can be obtained. The industrial-sized conventional boilers with capacities of less than 58.6 MW_t (200 x 10^6 Btu/hr thermal input) are generally of the spreader-stoker type, while the relatively bigger units are designed for pulverized-coal firing. Thus, depending on the boiler capacity, the combustion losses in industrial FBC boiler units will be equal to or greater than those in conventional coal-fired boiler plants.

Stack Losses

Stack losses consist of the unrecovered latent and sensible heat in the flue gas exiting through the stack. The stack losses depend on the stack gas temperature, excess air rate, and the coal composition, especially hydrogen and moisture content. The stack gas temperature is determined, in general, by the dew point of the flue gas, which is a function of the concentration of the SO_x and water vapor in the flue gas (sulfuric acid [H_2SO_4] mist formation tendency and plume buoyancy effects). In uncontrolled conventional coal-fired plants the stack inlet temperatures for flue gases from burning high-sulfur and low-sulfur coals are usually around 200°C (400°F) and 175°C (350°F), respectively.[35] In this study the stack inlet temperature for the industrial FBC boiler design has been assumed to be 135°C (275°F), and the fluidized-bed boilers have been designed to operate with 20 percent excess air.

As shown in Tables 18 and 19, the dry flue gas sensible heat losses are around 4.4 percent. The conventional industrial boilers are usually designed for excess air rates of about 50 percent for the stoker-fired and about 30 percent for the pulverized-coal-fired units. Thus, the relatively higher excess air rates and the stack inlet temperatures in conventional firing can result in dry flue gas sensible heat losses more than 50 percent greater than those in the FBC combustors.

The major portion of the energy loss due to hydrogen and moisture in coal is in the form of latent heat. The water vapor latent heat losses, being a major function of the coal composition, are comparable in conventional and FBC boiler plants. In industrial boilers using wet-scrubber FGD systems for SO_2 control, additional stack moisture losses will occur.

In pulverized-coal firing, and in some FBC designs with under-bed coal feeding, coal drying is required to limit the coal feed moisture content to about 5 percent. This can then result in additional energy losses through auxiliary power and fuel consumption by the drying system. For boiler efficiencies defined on the basis of using as-received coal, this additional energy requirement may have a minimal effect. In the case of high-moisture-content coals, however, like western subbituminous (20.8 wt % moisture), reducing the moisture level in the feed coal can result in increased energy available for steam generation in the bed and, hence, can affect the design of the in-bed heat transfer surface. The resulting reduction in the volumetric flow rate of hot flue gas, moreover, will affect the heat transfer surface requirements in the convective pass and in the economizer.

Radiation and Unaccounted Losses

Radiative heat losses depend on the surface emissivity, the fourth power of the absolute temperature, and the surface area of the boiler.

The unaccounted energy losses include convective losses occurring through the boiler walls, steam pipes, and other equipment at temperatures higher than ambient. The design philosophy in this study has been to maximize shop fabrication, and the relatively bigger plants [MW_t (150×10^6 Btu/hr thermal input) and more] have been designed to consist of multiple fluidized-bed combustor modules. Little difference may exist in the surface-to-volume ratios of the boilers designed to serve the four reference plant sizes. Thus, irrespective of the plant capacities, a constant value of 1.5 percent of the thermal input has been assumed to be lost as radiative and unaccounted for energy. In the case of conventional coal firing, the bigger plants (more than 22 MW_t) are designed for field-erected boilers, and, due to the possible reduction in surface-to-volume ratios, the radiation energy losses could show a dependence on the unit capacity.

Solids Sensible Heat

In this study we have assumed that the spent solids sensible heat is recovered. For the bituminous high-sulfur coal the maximum energy penalty of not recovering spent solids sensible heat could be around 1.0 to 1.5 percent, and for the bituminous low-sulfur and western subbituminous coal it could be less than 0.2 to 0.3 percent of the total thermal input. In conventional industrial boilers the rate of waste solids production could be significantly less than for the FBC boiler, but these solids may be leaving the boiler at a higher temperature than the waste solids in a FBC boiler. Little difference may exist, therefore, in the energy penalties due to solids sensible heat losses in conventional boilers and in industrial FBC boilers.

Calcination and Sulfation Energy Losses

The impact of SO_2 control on FBC boiler efficiency will be determined mostly by the calcination and sulfation energy requirements. For the bituminous high-sulfur coal the sorbent requirements for a given SO_2 emission control level are considerably higher than for the two low-sulfur reference coals and, therefore, result in sorbent calcination and sulfation energy losses of up to 2.2 percent for the 500 μm and up to 4.5 percent for the 1000 μm average bed particle size. In the case of bituminous low-sulfur and western subbituminous coals the analogous energy losses are of the order of only 0.4 to 0.8 and 0.1 to 0.2 percent, respectively. Note that the sulfation of the sorbent is an exothermic reaction providing a positive gain to the boiler efficiency that is independent of the sorbent feed rate.

Auxiliary Power Consumption

The draft losses and the auxiliary power consumption factors assumed in this study for estimating industrial FBC boiler plant auxiliary power requirements are given in Table 20. The total auxiliary power consumption, depending on the plant size, fluid-bed operating conditions, coal and sorbent characteristics, and the SO_2 emissions control standard, has been estimated to range from 3.0 to 3.9 percent of the thermal input. Fan power requirements are the major auxiliary power factor, and the effect of sorbent consumption rate on the total auxiliary power requirement is small. Thus, the NSPS for SO_2 has an insignificant effect on auxiliary power consumption.

While it is not exactly meaningful for industrial boiler applications, the auxiliary power electrical rate has been converted to thermal power by assuming that electrical power has been supplied at a 33 percent fuel-to-electricity efficiency. This provides, then, an estimate of an overall plant efficiency for the industrial FBC boiler plants. In the case of bituminous high-sulfur coal, even though the sorbent consumption with 500 μm particles is less than with 1000 μm particles, an additional 0.2 to 0.4 percent of thermal input could be needed as auxiliary power for the size reduction operation, depending on the availability of properly sized sorbent and the attrition behavior of the sorbent. Due to the relatively smaller amounts of sulfur in the two low-sulfur reference coals, no significant increase in the auxiliary power consumption for size reduction operation would be needed. The overall plant efficiency and boiler efficiency is sensitive to the NSPS for SO_2, the sorbent activity, and the type of coal, especially for the high-sulfur coal.

Table 20 Dwg. 7707A29

INDUSTRIAL FBC BOILER PLANT LOSSES AND POWER CONSUMPTION FACTORS

Component	Auxiliary Power Consumption Draft Losses, m (in) of Water
Forced Draft Ducts	0.005 (0.2)
Fluidized-Bed Chamber Section Dampers	0.020 (0.8)
Air Distributor Plate	0.305 (12.0)
Bed (1.22 m expanded bed depth)	1.057 (41.6)
Mechanical Particulate Collector	0.127 (5.0)
Convective Section	0.142 (5.6)
Economizer	0.051 (2.0)
Economizer Flues	0.005 (0.2)
Fabric Filter with Associated Flues	0.152 (6.0)
Total	1.864 (73.4)
FD and ID Fans (70% efficiency), kW/MW_t	8.14
Boiler Feedwater System,[a] kW/MW_t	2.26
Coal Handling System, kWh/Mg	2.75
Sorbent Handling System, kWh/Mg[b]	9.00
Waste Solids Handling System, kWh/Mg	Negligible

a Assumed to be a function of plant size only and independent of steam conditions.

b For 1000 μm average bed particle size.

The uncertainty in the assumed combustion efficiencies is greater than the impact of the Ca/S ratio on the overall plant efficiency. The sensitivity of the industrial FBC boiler efficiency and of the electricity utility AFBC boiler efficiency to the Ca/S ratio and coal properties are identical. The overall plant efficiency for the industrial FBC boiler (steam energy per unit of fuel energy input) is more sensitive to Ca/S ratio than is the electric utility AFBC power plant efficiency (electric power generated per unit of fuel energy input) because of the low efficiency of conversion of steam energy to electricity (∿41%).

Comparison of the Industrial FBC Boiler and Conventional Industrial Boilers

The boiler efficiencies and auxiliary power for the industrial FBC boiler and conventional coal-fired boiler systems are summarized in Table 21. These efficiencies for the conventional boilers with limestone wet-scrubbing have been estimated by assuming 50 percent excess air and 97 percent combustion efficiency for the spreader-stoker units, and 30 percent excess air and 99 percent combustion efficiency for the pulverized coal-fired units.[7] The 8.8 MW_t conventional coal-fired plant has been designed with a shop-fabricated boiler, whereas the design of the other three plants incorporates field-erected units. The radiative and unaccounted for energy losses for the conventional boilers have been assumed to decrease from 3.0 to 1.54 percent of thermal input as the plant size increases from 8.8 to 58.6 MW_t.[7] In the case of industrial FBC boiler systems, a modular and shop-fabricated FBC boiler design philosophy has been pursued. Because of relatively small differences in the surface-to-volume ratios of the boilers in the four FBC boiler plants, and also because of the relatively lower temperatures in the fluidized-bed combustion zone (840°C), a constant value of 1.5 percent has been assumed to represent the radiative and unaccounted for energy losses. The stack losses for the conventional boilers have been estimated for two cases: no reheat of the stack gas and reheat to 79°C.

The energy requirements for the conventional industrial boiler flue gas desulfurization system have been reported to be a function of coal

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Table 21

COMPARISON OF BASE INDUSTRIAL FBC BOILER AND UNCONTROLLED CONVENTIONAL BOILER ENERGY EFFICIENCIES[a]

Coal Type	Plant Size, MW_t	Conventional Boiler Type.	Boiler Efficiency, %			Auxiliary Power, kW_e	
			Conventional with Limestone FGD System[a]		FBC[b]		
			Without Reheat	With Reheat		Conventional[a, c]	FBC[d]
Bituminous High - Sulfur, 3.5 wt % S	8.8	Spreader-Stoker	79.17	77.44	82.21 - 84.3	104-123	100-115
	22.0	Spreader-Stoker	80.01	78.27		244-290	249-287
	44.0	Spreader-Stoker	80.60	78.86		476-569	499-574
	58.6	Pulverized	84.15	82.64		605-713	665-766
Bituminous Low - Sulfur, 0.9 wt % S	8.8	Spreader-Stoker	81.7	79.94	84.96-85.34	100-105	91-94
	22.0	Spreader-Stoker	82.54	80.77		234-247	228-235
	44.0	Spreader-Stoker	83.08	81.31		456-486	456-470
	58.6	Pulverized	86.61	85.07		570-609	608-626
Western Sub-bituminous, 0.6 wt % S	8.8	Spreader-Stoker	79.35	77.60	83.12-83.35	100-105	92-95
	22.0	Spreader-Stoker	80.19	78.42		234-247	233-237
	44.0	Spreader-Stoker	80.74	78.98		456-486	461-475
	58.6	Pulverized	83.27	81.73		579-609	615-633

[a] Controlled conventional boiler information from References 28 and 31.

[b] Range for the three reference sorbents (500 μm particle size) ; lower limit - low-activity sorbent (Menlo Quarry Stone) at stringent control; upper limit - high activity sorbent (Western 90% CaL Stone) at moderate control.

[c] Lower limit - SO_2 emissions moderate control; upper limit - SO_2 emissions stringent control.

[d] Range for the three reference sorbents (500μm particle size) ; lower limit - high-activity sorbent (Western 90% CaL Stone) at moderate control; Upper limit - low-activity sorbent (Menlo Quarry Stone) at moderate control.

sulfur content, SO_2 control level, and, to a lesser extent, plant size.[8] For the lime/limestone-based FGD systems the energy requirements for the three reference coals and the SO_2 emission control levels have been estimated to range from 0.9 to 1.8 percent without exhaust gas reheat and 2.4 to 3.5 percent with exhaust gas reheat.[36] These energy requirements are expressed as a percent of boiler thermal input and are comparable to the sulfation and calcination energy losses in industrial FBC boilers.

FBC boiler stack losses, however, irrespective of coal type, have been estimated by assuming a stack inlet temperature of 135°C (275°C). Thus, accounting for factors such as combustion efficiencies (FBC boiler - 95%; spreader-stoker - 97%; pulverized type - 99%), stack losses (function of excess air rate and stack inlet temperature), radiative and unaccounted for energy losses (1.5% for all FBC boiler plants; 3.0% for 8.8 MW_t, 2.2% for 22.0 MW_t, 1.7% for 44.0 MW_t, and 1.54% for 58.6 MW_t conventional plants[4]), and calcination and sulfation energy requirements (function of sorbent reactivity and SO_2 control standard for the FBC boiler), the FBC boiler efficiencies summarized in Table 21 are higher than those for spreader-stoker units but are generally lower than those for the pulverized coal-fired boiler without reheat. It is the higher combustion efficiency of the pulverized coal-fired unit that is mainly responsible for its better performance in this case. The conventional boiler efficiency is very sensitive to the stack gas reheat temperature, dropping below the FBC boiler at 79°C.

In conventional coal-fired industrial boilers, even though relatively higher excess air rates are used, the fan power consumption is less, because of lower draft losses. Overall, the auxiliary power consumption for the conventional and FBC industrial boilers is comparable.

PARAMETRIC PERFORMANCE

Fluidized-bed design and operating conditions, such as the fines recycle option, coal feed location, excess air rate, gas velocity, bed depth, and the SO_2 control standard, will affect the energy comsumption of the industrial FBC boiler system. The effect on FBC boiler energy performance due to variations in these parameters are discussed below.

Recycle vs. Nonrecycle of Primary Cyclone Fines to the Combustor

The fines elutriated from the combustor bed consist of unburned char, coal ash, and partially sulfated sorbent particles. For improved carbon utilization these fines can be collected and burned in a separate carbon burnup cell or reinjected into the combustion bed. The choice between the two will be determined by the economics as well as by the reliability and availability of the industrial FBC boiler system. In the case of electric-utility-sized FBC boilers, due to the relatively larger throughputs involved and the higher combustion efficiencies desired, FBC design concepts incorporating a separate carbon burnup cell have been favored. For economic reasons, however, recycle of collected carry-over fines into the combustor bed may be desirable for the industrial-sized FBC boiler. As mentioned previously, combustion efficiencies of 85 to 96 percent could be expected for once-through carbon operation, and, with the recycle of carry-over fines to the combustor bed, the carbon utilization could be improved to give as high as 95 to 97 percent combustion efficiency. In this study the reference FBC design incorporates the recycle option and has assumed a combustion efficiency of 95 percent for determining energy impact. Thus, operating the reference FBC boiler without recycle could result in additional energy losses of up to 10 percent.

Coal Feed Location

Fluid-bed boiler designs proposed include both in-bed and above-bed coal feeding. The base design in this study assumes in-bed feeding because above-bed feeding may result in incomplete combustion in the bed and freeboard combustion of volatiles, thus affecting SO_2 capture. The potential problem of incomplete combustion could be handled by reinjecting the collected carry-over material into the combustor bed, by using relatively higher excess air rates and larger freeboard volume to facilitate complete combustion of volatiles in the freeboard, and/or by feeding carefully sized (or wet) coal. These options, however, could penalize the boiler efficiency through increased stack losses and could still result in increased SO_2 emissions.

Impact of Industrial FBC Boiler Operating Conditions

Excess air rate, bed depth, gas velocity, and Ca/S molar ratio either individually or in combination with each other can dramatically affect fluidized-bed performance. The effect of these variables on FBC boiler and plant efficiencies has been examined for the eastern high-sulfur coal using 1000 µm average bed particle size and high-activity sorbent (Western 90% CaL stone) to meet the stringent SO_2 emissions control level. The results are summarized in Table 22. When the excess air rate is increased from 20 percent to 100 percent, there is an increase of about 2.1 percent in stack losses and about 1.7 percent in auxiliary power consumption, which severely affects FBC boiler performance. This could be countered, though, by a significant increase in combustion efficiency. Operating the fluidized-bed at higher velocities results in shortened gas residence times and, therefore, will require more sorbent for meeting the SO_2 emissions control level, which will result in increased energy penalties, mostly through sorbent calcination energy requirements. One of the options for minimizing the sorbent consumption rate and increasing combustion efficiency is to operate with deeper beds. The auxiliary power increases from 3.18 to 3.95 percent when the bed depth is increased from 1.22 to 1.83 m. In the particular case examined, because of the high reactivity of the sorbent, no significant reduction in sorbent consumption is realized by increasing bed depth. Also, we have assumed no increase in combustion efficiency. With poor reactivity sorbents, however, there may be enough reduction in sorbent consumption at least to nullify the increased auxiliary power requirements incurred by increasing the bed depth.

Table 22 Dwg. 7707A31

IMPACT OF FLUIDIZED-BED OPERATING CONDITIONS OF FBC BOILER AND PLANT EFFICIENCIES[a]

	Base Case	Parametric Assessment		
	Excess Air, 20%; 1. 83 m/ s; 1. 22m	Excess Air, 10-100%	Superficial Velocity, 1. 22-3. 05 m/ s	Expanded Bed Depth 0. 91-1. 83 m
Ca/S Molar Ratio	3. 45	3. 45	3. 34- 5. 29	4. 30-3. 29
Energy Losses[b]				
Unburned carbon	5. 00	5. 00	5. 00	5. 00
Stack losses	9. 11	8. 80-11. 20	9. 11	9. 11
Radiation and unaccounted	1. 50	1. 50	1. 50	1. 50
Calcination and sulfation	0. 92	0. 92	0. 84- 2. 31	1. 56- 0. 80
Total Losses	16. 53	16. 22-18. 62	16. 45-17. 92	17. 17-16. 41
Boiler Efficiency[b]	83. 47	83. 78-81. 38	83. 55-82. 08	82. 83-83. 59
Auxiliary Power[b]	3. 18	3. 00 - 4. 70	3. 15 - 3. 45	2. 90 - 3. 95
Plant Efficiency[b]	80. 29	80. 78 -76. 68	80. 40 -78. 63	79. 93 -79. 64

[a] Bituminous high-sulfur coal, 1000μm average bed particle diameter, high-activity sorbent (Western 90 % CaL stone), 90 % sulfur removal

[b] Expressed as percent of thermal input.

CONCLUSIONS

For industrial FBC boiler systems the energy conversion efficiency depends strongly on the coal sulfur content, sorbent reactivity, fluidized-bed operating conditions, and, to a lesser extent, the SO_2 emission control standard. For a given set of fluidized-bed design and operating parameters, these energy effects would be felt mostly through the calcination energy loss. There may also be some relatively minor variations in the auxiliary power requirements of the sorbent handling system. Depending on the sorbent activity category, the calcination and sulfation energy losses (with 500 μm average bed particle diameter to meet the stringent SO_2 emission control level) for bituminous high-sulfur coal range from 0.45 to 2.18 percent, and from 0.04 to 0.39 percent for bituminous low-sulfur and western subbituminous coals. In the case of fluidized-bed operation with 1000 μm average bed particle diameter, similar energy losses for bituminous high-sulfur coal have been found to range from 0.92 to 4.55 percent and, for the bituminous low-sulfur and western subbituminous coal, from 0.13 to 0.82 percent. On changing the SO_2 emissions control level from moderate to stringent, and depending on the sorbent activity category and particle size, the boiler efficiencies for the bituminous high-sulfur coal vary by 0.6 to 1.5 percent and, for the two low-sulfur coals, by 0.05 to 0.16 percent. The industrial FBC boiler compares favorably with the conventional industrial boiler with respect to energy conversion efficiency, and the ACL for SO_2 does little to affect this comparison.

12. COST ASSESSMENT

Industrial FBC boiler systems in the capacity range covered in this study have not yet been demonstrated on a commercial scale. Because of the developmental nature of FBC technology, comprehensive cost information is as yet unavailable. In this section cost projections developed to assess the impact of alternative emission control levels are presented for the base industrial FBC boiler design and discussed parametrically for key design and operating variables. The cost projections reflect trends in cost sensitivity, but the absolute costs are, by their nature, limited in accuracy and, hence, should be used with caution.

COST PREMISES

The costing premises used in this study are summarized in Tables 23 and 24. They were compiled by EPA for use in all their industrial boiler studies to promote consistency. The industrial FBC boiler costs are presented for grass roots plants in the Midwest. The plant boundaries include inlets to coal and sorbent storage, boiler feedwater inlet to the economizer, steam outlets from the steam generator, on-site spent solids storage outlets, and the stack outlet. The costs for the steam and condensate return lines from the process area are not included.

The bases for costs developed are presented in Appendix D.

BASE INDUSTRIAL FBC BOILER PLANT COSTS

The total capital investment for industrial FBC boilers meeting the stringent SO_2 control levels is compared with that for conventional boilers with limestone FGD system in Figure 30. The uncontrolled, conventional boiler cost estimates developed by PedCo[37] and the limestone FGD system costs developed by the Radian Corporation have been used to generate the controlled conventional boiler cost data.[36] The results in Figure 30 show that for the bituminous coals the industrial FBC boiler

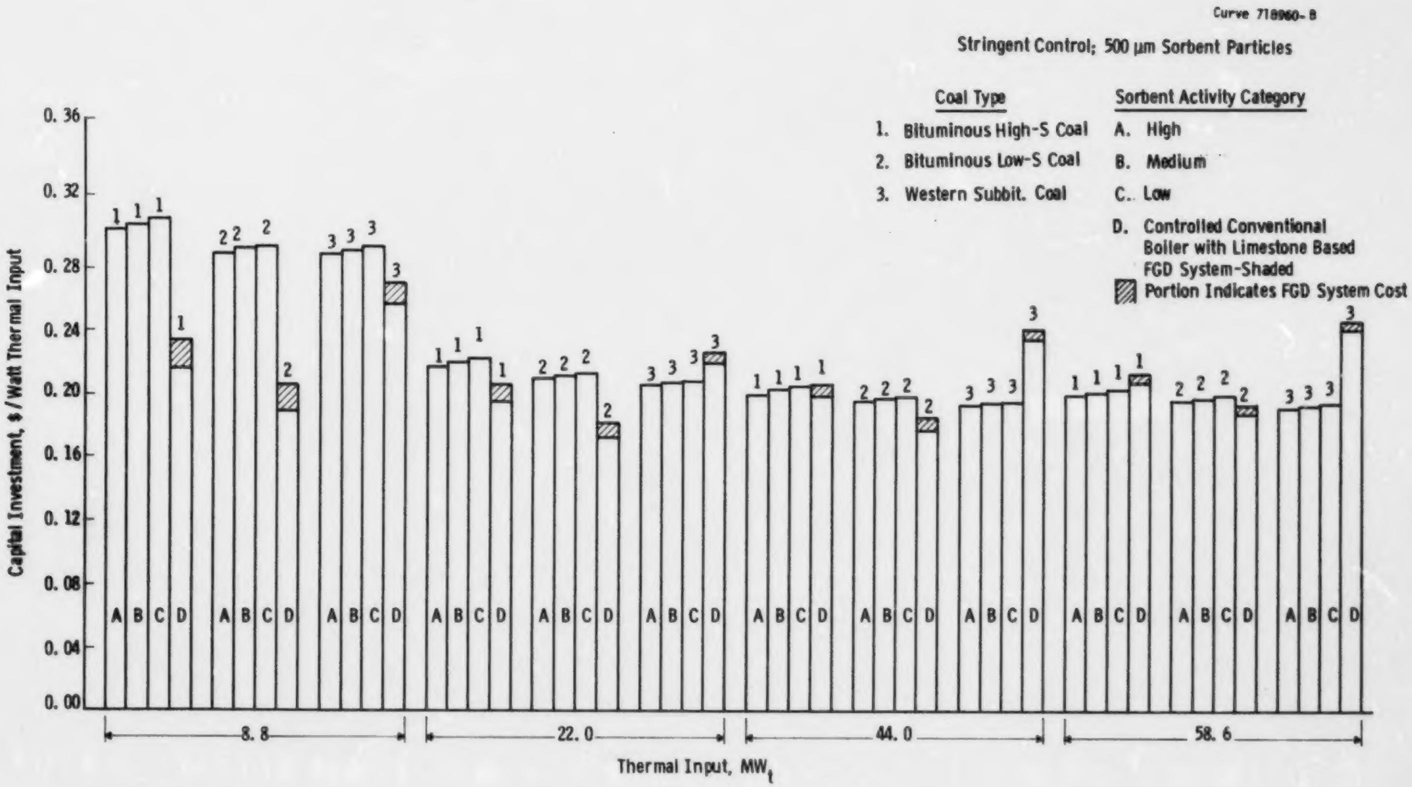

Figure 30 - Capital Costs for Industrial FBC Boiler Plants Compared with Uncontrolled and Controlled Conventional Industrial Boiler Plants

Table 23

COST FACTORS FOR ESTIMATING CAPITAL COSTS

Cost Base Year	Mid-1978
Site	Midwest
Direct Costs (DC)	Equipment cost + installation cost = installed cost
Indirect Costs (IC):	
Engineering services	10% of DC
Construction and field expenses	10% of DC
Contractor's fee	10% of DC
Start-up	2% of DC
Contingencies	20% of (DC + IC)
Turnkey Cost (TC)	DC + IC + contingencies
Working Capital (WC)	25% of total direct operating cost
Total Capital Investment	TC + WC

capital investment for units of 44.0 MW_t is comparable to that for the conventional boilers with limestone wet-scrubbing. For the western subbituminous coals the industrial FBC boilers have a capital cost advantage over the conventional industrial boilers for units of 22.0 MW_t or larger. The relatively lower ash-softening temperatures for the western coals are partially responsible for the higher costs of the conventional boiler with western coal, whereas, due to the relatively lower FBC operating temperatures, the ash fusion points will have little effect on the FBC boiler costs. The coal type, sorbent reactivity, or, therefore, the SO_2 standard have little impact on the capital investment per unit of thermal energy input (see Appendix D). Two other recent cost studies have shown similar capital cost comparisons of industrial AFBC and conventional boilers.[38,39]

38,39

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Table 24

COST FACTORS FOR ESTIMATING ANNUAL COSTS[a]

Plant Capacity Factor, %	60
Direct Operating Cost (DOC):	
Direct labor, \$ / man-hour	12.02[b]
Supervision, \$ / man-hour	15.63[c]
Maintenance labor, \$ / man-hour	14.63[b]
Electricity, mills/ kWh	25.8[d]
Replacement parts (includes maintenance materials)	
Untreated water, \$ / 1000 gal.	0.12
Process water, \$ / 1000 gal.	0.15
Boiler feedwater, \$ / 1000 gal.	1.00
Coal, high S, \$ / 10^6 Btu (\$ / ton) (eastern)	0.74 (17.00)[e]
low S, \$ / 10^6 Btu (\$ / ton) (eastern)	1.16 (29.00)[e]
low S, \$ / 10^6 Btu (\$ / ton) (Wyoming)	0.42 (6.75)[e]
Limestone, \$ / ton delivered (Western 90% CaL)	25.00
\$ / ton delivered (Bussen quarry)	25.00
\$ / ton delivered (Menlo quarry)	25.00
Sodium hydroxide, \$ / ton (bulk, 50% basis, FOB works)	158.00[f]
Sodium carbonate, \$ / ton (bulk, FOB works)	60.00[f]
Ammonia, \$ / ton delivered	130.00[f]
Ammonium hydroxide, \$ / ton (29.3% NH_3 basis, freight equalized)	173.00[f]
Solid waste disposal, \$ / ton	8.00
Overhead Cost (OC):	
Payroll	30% of direct labor
Plant	26% of labor, parts and maintenance
Capital Charges (CC):	
G & A, taxes and insurance	4% of turnkey costs
Capital recovery	10.6% of turnkey costs
Interest on working capital where working capital is 25% of DOC	10% of working capital
Total Annual Costs (TAC)	DOC + OC + CC

a All costs are in June 1978 dollars.

b Engineering News-Record, June 29, 1978, pp. 52-53, average for Chicago, Cincinnati, Cleveland, Detroit, and St. Louis.

c Estimated at 30 percent over direct labor rate.

d EEI members publication for June 1978, average for Boston, Chicago, Indianapolis, Houston, San Francisco, and Los Angeles.

e Coal Outlook, July 18, 1978, issue, spot market prices.

f Chemical Marketing Reporter, June 19, 1978.

The annual cost estimates shown in Figure 30 favor industrial FBC boilers over the controlled conventional industrial boilers. Radian Corporation has reported that for the smaller boilers (less than 22 MW_t), the sodium throwaway process has the lowest capital and annual cost requirements.[36] Even for conventional industrial boilers with sodium throwaway FGD systems, the annual costs are more than for the industrial FBC boiler. The operating labor and materials costs in the industrial FBC boiler system are similar to those in the uncontrolled conventional industrial boilers. This similarity could be due to the integrated nature of sulfur removal and steam generation processes in the AFBC system. In the controlled conventional boiler plants, however, additional operating labor and material costs for the FGD system are incurred. Radian Corporation has reported that depending upon the boiler size, fuel sulfur content, SO_2 control level, and type of FGD system (limestone, sodium throwaway, double alkali, Wellman-Lord, and spray-drying), the additional capital and annual cost requirements could be 10 to 70 percent and 15 to 50 percent, respectively, of the uncontrolled conventional industrial boiler costs. Obviously, there is a great deal of uncertainty in the absolute costs, and the costs are very sensitive to the cost premises applied.

The annual costs defined per unit of energy output ($/watt thermal output) for the three reference coals are plotted in Figure 31. In comparison with western subbituminous coal (0.6 wt % S), the relatively higher industrial FBC boiler annual costs for the bituminous low-sulfur coal (0.9 wt % S) is due mostly to the higher coal cost, whereas for the bituminous high-sulfur coal (3.5 wt % S) the contribution of the sorbent costs is considerable.

The steam costs ($/Mg) for the industrial FBC boiler are shown in Figure 32. The effect of plant capacity, coal type, sorbent activity, sorbent particle size, and the ACL for SO_2 is indicated in the figure. These parameters have a greater effect on the bituminous, high-sulfur coal case; the SO_2 standard is of secondary importance. The costs presented here are in agreement with the trends indicated by two other recent industrial FBC cost studies.[38,39]

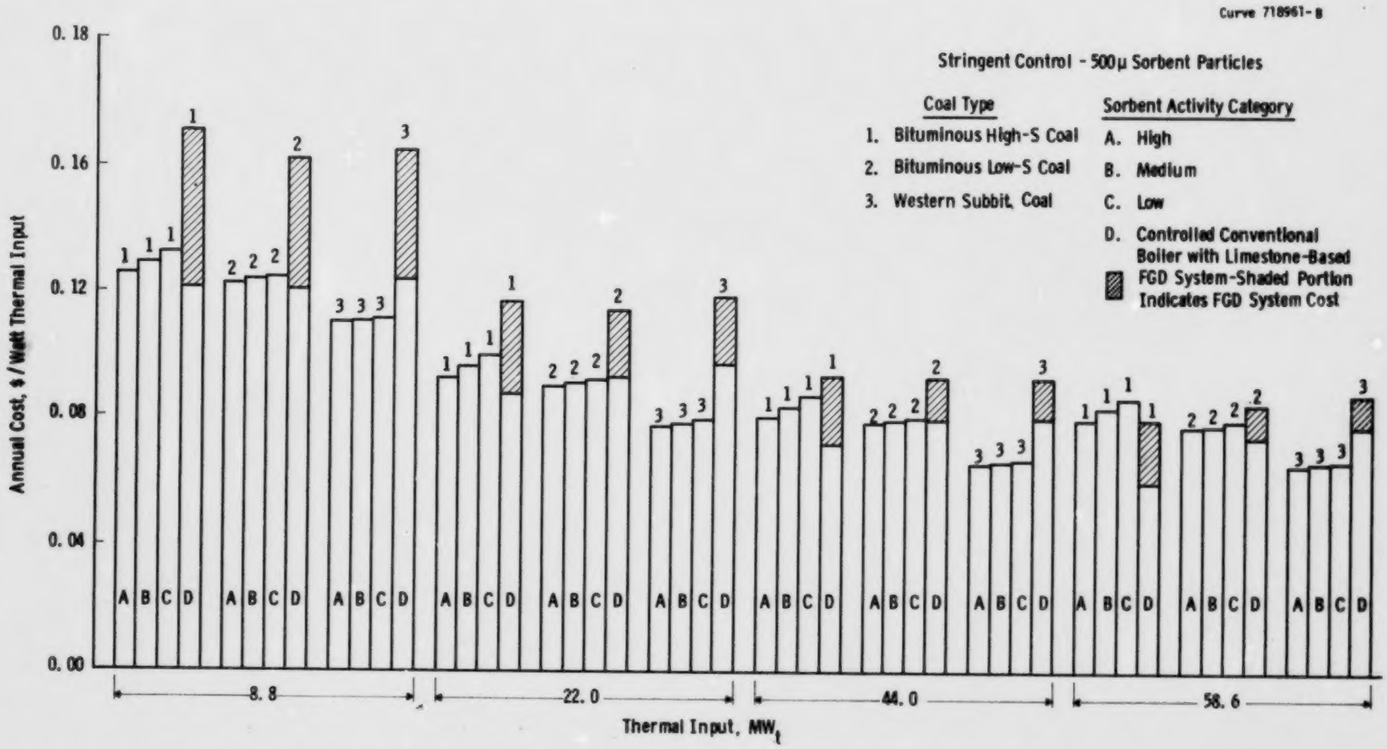

Figure 31 - Annual Costs for Industrial FBC Boiler Plants Compared with Uncontrolled and Controlled Conventional Industrial Boiler Plants

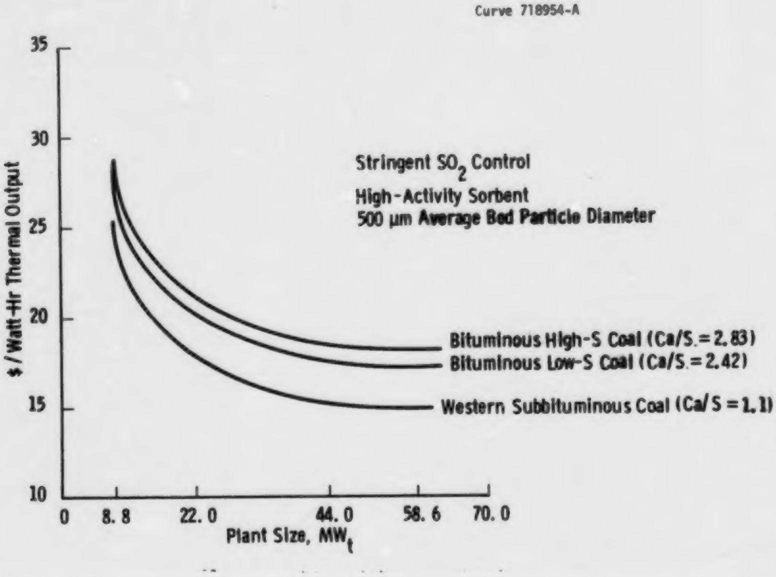

Figure 32 - Industrial FBC Boiler Annual Costs

Effect of Steam Conditions on the Base Boiler Design Costs

The steam production rates (kg/hr) per unit plant thermal output will vary as a function of the steam conditions and, hence, will result in lower or higher steam costs under similar fluidized-bed operating conditions and boiler and plant efficiencies. The 8.8 and 22 MW_t plants generate 1135 kPa (150 psig) saturated steam, whereas the 44 and 58.6 MW_t plants generate 3204 kPa, 316°C (450 psig, 600°F) and 5272 kPa, 399°C (750 psig, 750°F) superheated steam. The 8.8 and 22.0 MW_t plants contain one module each, whereas the 44 and the 58.6 MW_t plants contain two and three boiler modules, respectively. Thus, the relatively higher steam costs for the 58.6 MW_t plant is due both to higher capital investment (three boiler modules) and to relatively lower steam throughputs [(kg steam/hr)/W thermal output)].

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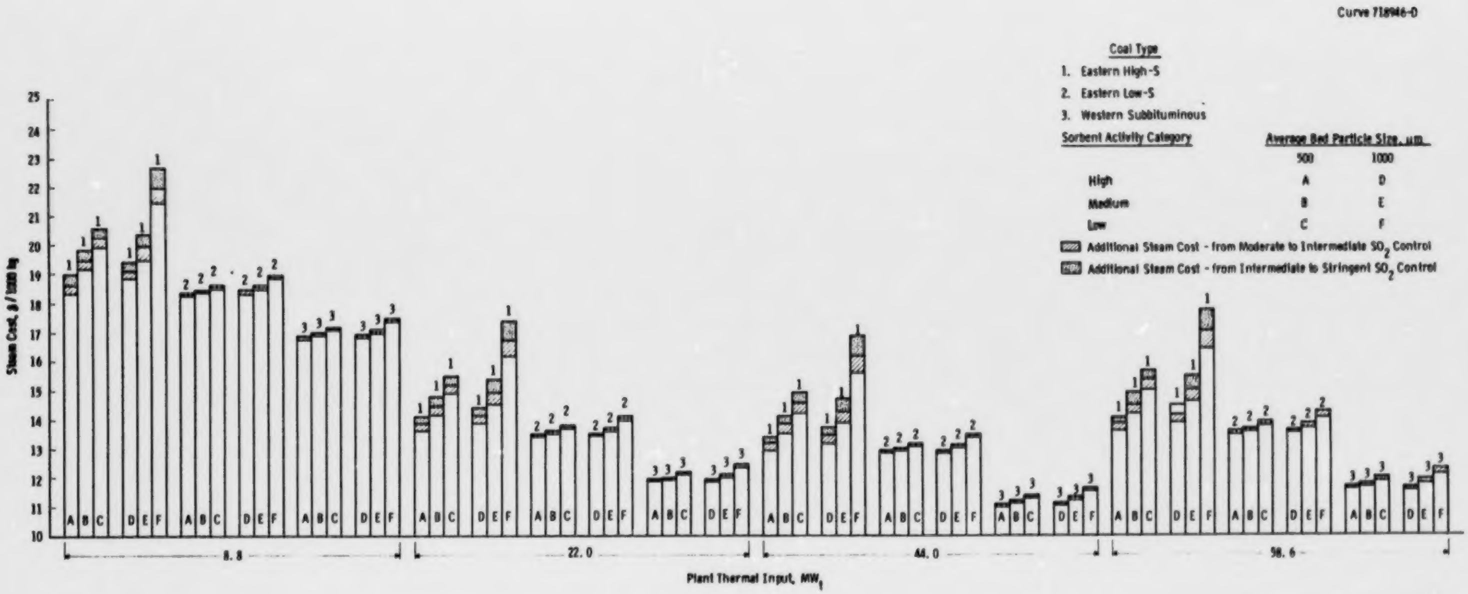

Figure 33 - Industrial FBC Boiler Plant Steam Cost

Effect of Coal Cost and Sorbent Consumption

The steam costs have been calculated by assuming the bituminous high-sulfur, bituminous low-sulfur, and western subbituminous coal costs, expressed as $\$/MW_t$, to be 2.523, 3.955, and 1.432, respectively. The sulfur content in the three reference coals, bituminous high-sulfur, bituminous low-sulfur, and western subbituminous coals, expressed as ng sulfur/J (lb sulfur/10^6 Btu), are 1275, 280, and 269 (2.97, 0.65, and 0.63), respectively. In comparison with those for western subbituminous coals, the higher steam costs for the bituminous low-sulfur coal is due to the higher coal cost. Thus, for coals with similar sulfur contents, a difference of about \$2.5 MW_t in coal cost results in an increase in steam cost of about \$1.5 to 2.0/Mg of steam produced. Of course, the exact number will depend on the coal sulfur content, SO_2 control level, plant size, and steam conditions. Increasing the coal costs by about 175 percent, therefore, could result in about an 8-to-17 percent increase in steam costs.

The relatively higher steam costs for the bituminous high-sulfur coal are due to higher sorbent consumption rates. In this study sorbent cost has been assumed to be \$28/Mg (\$25/ton), irrespective of the limestone source. To meet the stringent SO_2 control level while burning bituminous high-sulfur coal, the Ca/S molar ratio requirements (using 500 μm average bed particle sizes) for the high-reactivity and low-reactivity sorbents are 2.83 and 5.26, respectively. Thus, as the sorbent consumption is increased by about 85 percent, the steam costs increase by about 8 to 12 percent. The Ca/S molar ratio requirement, while using 1000 μm average bed particle sizes for the low-reactivity sorbent, is 8.46. With Ca/S molar ratios of 2.83 and 8.46, there is a 20 to 26 percent increase in the steam cost when the sorbent consumption is increased by about 300 percent.

500 μm

Effect of SO_2 Control Level on the Base Boiler Design Cost

The ratio of steam costs at stringent control to those at the moderate control level have been calculated. The results are summarized in Table 25. These ratios have been calculated by dividing the steam cost for a given coal and sorbent type at the stringent control level into that at the moderate control level. These results indicate little impact of SO_2 control level on the steam costs for the bituminous low-sulfur and western subbituminous coals. For the bituminous high-sulfur coal, however, the steam cost increases in the range of 4 to 9 percent when the SO_2 emission standard is changed from the moderate to the stringent control level. If no sulfur removal were required, in the extreme case, the cost of steam would be reduced significantly for the bituminous high-sulfur coal ($0.5 to 1.5/1000 lb), but the reduction would be negligible for the low-sulfur coals.

Table 25

RATIO OF STEAM COSTS AT STRINGENT CONTROL TO THOSE AT MODERATE CONTROL[a]

Dwg. 7707A30

Sorbent Activity Category	High		Medium		Low	
Average Bed Particle Size, μm	500	1000	500	1000	500	1000
Coal Type						
Bituminous high-sulfur	≤1.04	≤1.05	≤1.05	≤1.06	≤1.05	≤1.09
Bituminous low-sulfur	≤1.005	≤1.005	≤1.005	≤1.005	≤1.006	≤1.009
Western subbituminous	≤1.005	≤1.005	≤1.006	≤1.006	≤1.006	≤1.010

[a] For the four reference plant sizes.

PARAMETRIC COST ASSESSMENT

Table 26 summarizes the relation between the study parameters and the various components of the plant. These variables, either individually or in combination with each other, could influence the industrial FBC boiler plant cost. These parameters, and the impact of their cost, are discussed in this section.

Effect of Steam Conditions

The steam temperature and pressure dictates the heat transfer surface area and materials requirements. The calcination energy losses influence the thermal energy available for steam generation in the bed and, hence, the submerged heat transfer surface requirement. Thus, SO_2 control level indirectly coupled with the plant steam generation requirements could affect the industrial FBC boiler cost. The installed boiler costs plotted in Figure 33 show that the capital requirements for producing superheated steam (5272 kPa, 750°C) could be 20 to 30 percent higher than those for saturated steam (1135 kPa) [150 psig] generation. The smaller temperature differential available as the driving force in the production of superheated steam results in increased heat transfer surface requirements and, hence, in increased boiler costs. The installed base boiler costs in the size ranges considered in this study are 20 to 25 percent of the total plant capital investment. Producing superheated steam instead of the saturated steam would thus raise the total plant investment by less than 10 percent. As shown in Figure 34, however, the steam conditions have a more significant impact on the steam costs than does the SO_2 control level. For a given plant size, generation of superheated steam results in relatively lower steam throughputs (kg/hr) and, hence, increased steam costs. Nevertheless, the annual costs are not significantly different; changes in steam conditions result in higher steam costs primarily because of changes in steam throughputs (kg/hr).

Table 26

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RELATION BETWEEN STUDY VARIABLES AND FLUIDIZED-BED BOILER PLANT COMPONENTS

Fluidized-Bed Boiler Plant	Boiler					Auxiliary			
Components / Study Parameters	Steam/Water Circuitry & Tubing Layout	Air Plenum, Grid Plate Design, Air/Flue Gas Ducting	Coal Feed System	Sorbent Feed System	Solids Withdrawal and Cooling System	Hot Gas Cleaning System	Solid Wastes Handling System	Coal Handling System	Sorbent Handling System
Steam Conditions	X								
Coal Characteristics	X		X		X	X	X	X	
Sorbent Type and Particle Size				X	X	X	X		X
Excess Air		X							
Superficial Velocity	X	X				X			
Bed Depth	X	X			X	X			
Bed Temperature	X	X			X				
Ca/S Molar Ratio				X	X	X	X		
Heat Transfer Coefficient	X								
Feed Points/Bed Area		X	X	X					
Bed Material Drain Points/Bed Area		X			X				
Control of Carbon Utilization						X	X		

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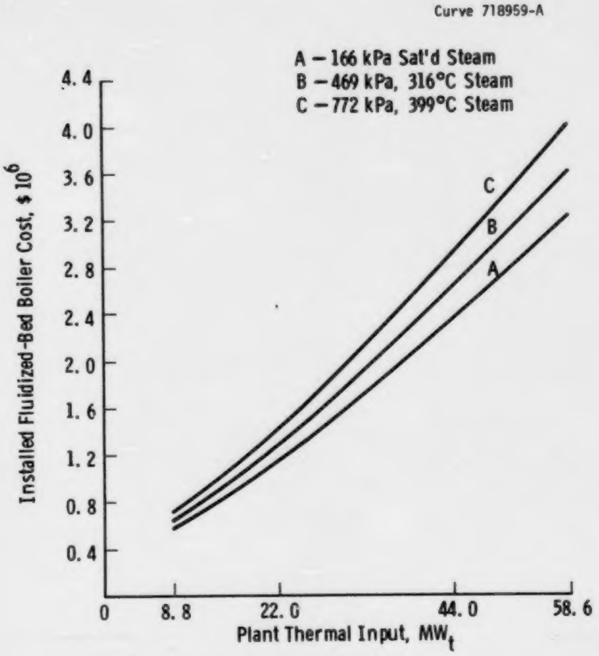

Figure 34 - Effect of Steam Conditions on Boiler Cost

The amount of steam generated depends on the energy available (boiler efficiency) and the enthalpy differential existing between the boiler feedwater inlet and steam generator outlet points. In this study the boiler feedwater has been assumed to enter the economizer at 93°C (200°F). Less than 100 percent return of the condensate, however, may require makeup water that could be at a significantly lower temperature. In the extreme case, where the boiler feedwater consists entirely of makeup water at 27°C (80°C), the steam production rates would be reduced by as much as 10 to 15 percent. The results tabulated in Table 27 show that under certain conditions the impact of the SO_2 control level could be less than that due to the boiler feedwater conditions at the inlet to the economizer. The boiler feedwater conditions, in practice, depend upon the specific requirements of the industrial steam application.

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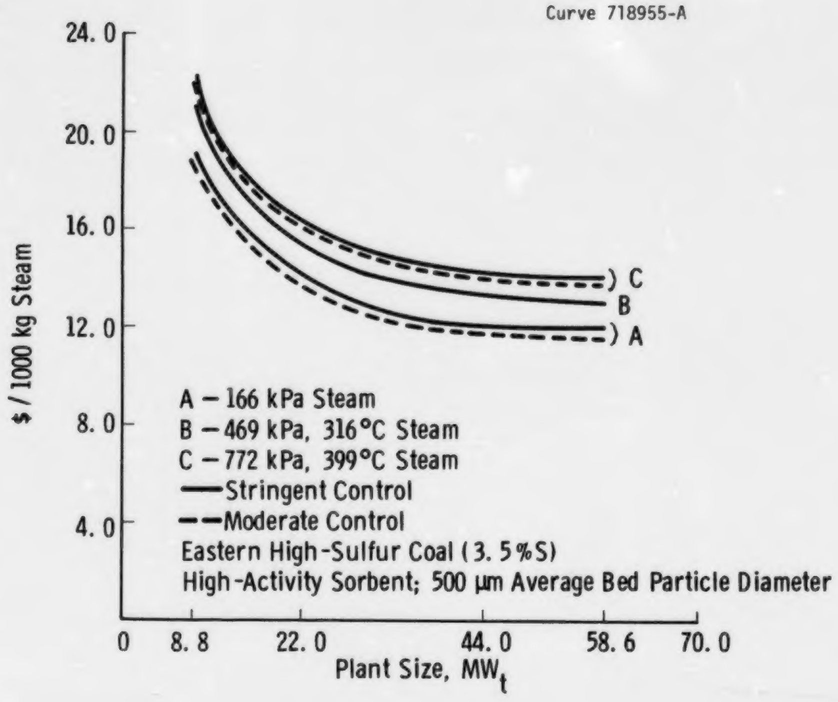

Figure 35 - Impact of SO_2 Control Standard on Steam Costs

Effect of Fluidized-Bed Operating Velocity

The base industrial FBC boiler design in this study assumes a gas fluidizing velocity of 1.83 m/s. The FBC boiler dimensions (cross-section) are determined by the gas velocity. The results plotted in Figure 35 show that operating velocity has little impact on FBC boiler cost, which itself is a small portion of the total plant cost. The design philosophy in this study ~~has~~ been to maximize shop fabrication; owing to the shipping constraints, therefore, the number of modules required is also influenced by the velocity.

The operating velocity will also affect the gas residence time and, hence, the sorbent consumption (calcination energy losses) and will, in turn, indirectly affect the steam costs. In most cases the sorbent consumption rate and its associated cost will dominate over the boiler

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Table 27

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EFFECT OF BOILER FEEDWATER INLET TEMPERATURE ON STEAM COSTS, $/1000 kg

Coal Type	Eastern high-sulfur (3.5 wt. % S)				
Sorbent Activity Category	High				
Sorbent Particle Size, µm	500				
SO_2 Control Level	Stringent	Moderate			
Boiler Feedwater Inlet Temperature, °C (°F)	93 (200)	93 (200)			
Makeup Water Temperature, °C (°F)	27 (80)	27 (80)			
Makeup Water, % of Boiler Feedwater	0	0	10	20	100
Plant Size (steam conditions)					
8.8 MW_t (166 kPa)	19.07	18.61	18.83	19.05	20.79
22.0 MW_t (166 kPa)	14.07	13.66	13.82	13.98	15.26
44.0 MW_t (469 kPa, 316°C)	13.41	12.97	13.11	13.25	14.36
58.6 MW_t (772 kPa, 399°C)	14.07	13.61	13.75	13.88	14.97

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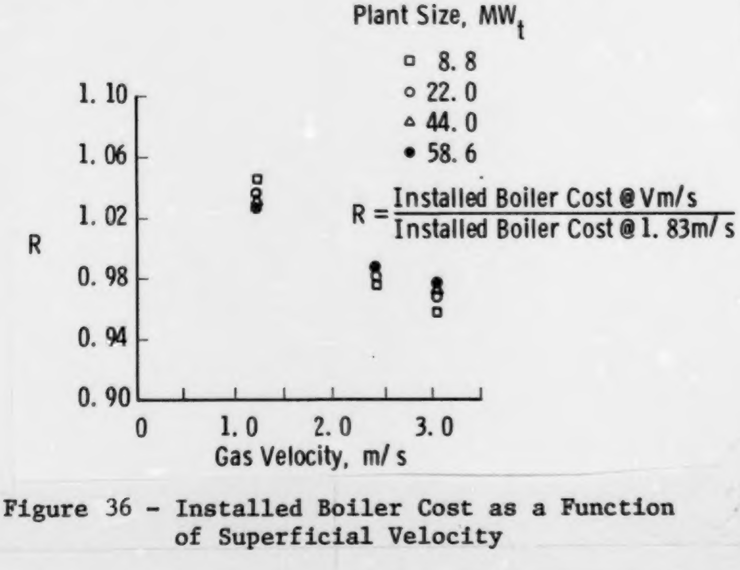

Figure 36 - Installed Boiler Cost as a Function of Superficial Velocity

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cross-section and its associated cost. In the lower-sulfur coal cases more flexibility is available in the selection of higher velocities because of the relatively smaller sorbent consumption characteristic of these coals. This is illustrated in Figures 35 and 36.

The particle size distribution and fluidized-bed operating velocity have a secondary effect on the heat transfer coefficient in the bed and, hence, the boiler cost. The heat transfer surface cost in terms of the overall plant capital investment, however, is rather small, and the impact on capital and annual costs is limited.

Sensitivity of Steam Cost to Boiler Efficiency

The boiler efficiency accounts for the unburned combustible losses, stack losses (flue gas flow rate [excess air], stack inlet temperature), calcination and sulfation energy requirements (sorbent activity category, sorbent consumption, SO_2 control standard), and radiation and unaccounted for losses. The ratio of steam costs plotted in Figure 37 shows that for every 1 percent change in boiler efficiency, the steam costs could change by approximately 1.2 to 1.5 percent.

Fluidized-bed boiler parameters such as the number of feed points, bed material drain points, carry-over fines reinjectors and their location, and bed operating temperature may have little effect on the plant total capital investment. These parameters, however, could affect the combustion and sulfur removal performance in the bed. Nonuniform combustion resulting in increased combustible and sorbent calcination energy losses could affect the steam costs through decreased combustion efficiency and/or increased sorbent consumption.

Impact of NO_x and Particulate Control on Steam Costs

In this study no specific costs for NO_x control have been added. If combustion modifications (staged combustion) and/or flue gas treatment are considered necessary to meet NO_x emission control standards,

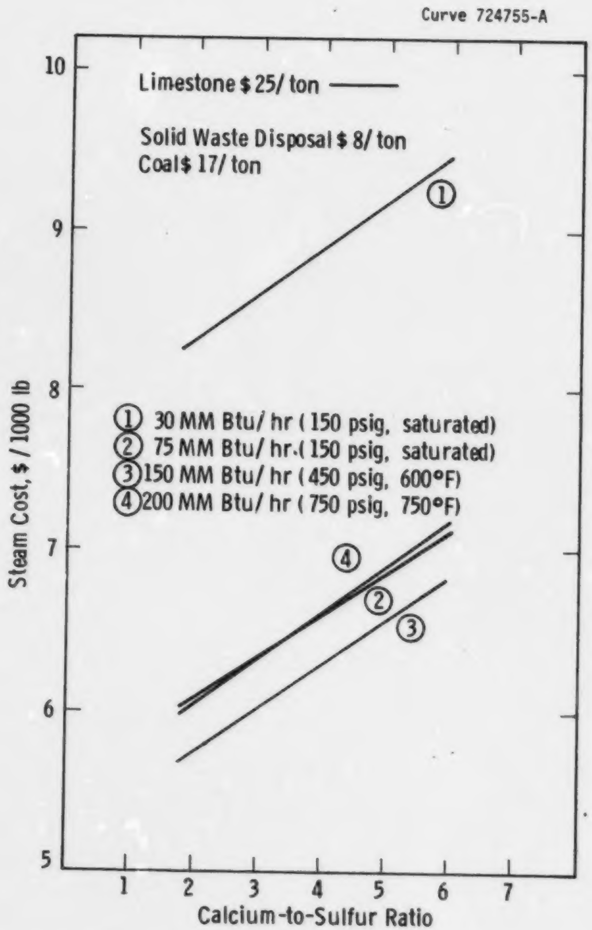

Figure 37 - Impact of Ca/S Ratio on Steam Cost with High-Sulfur Bituminous Coal

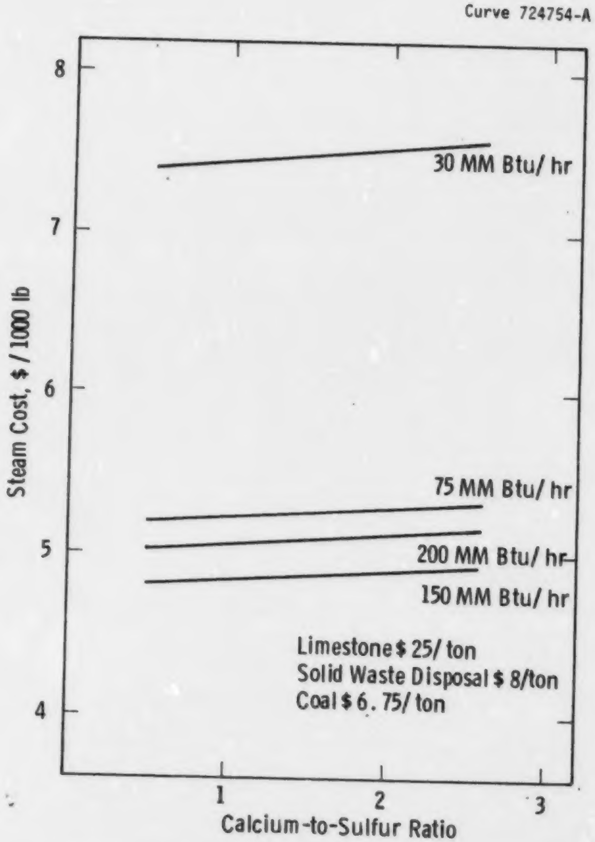

Figure 38 - Impact of Ca/S Ratio on Steam Cost with Subbituminous Coal

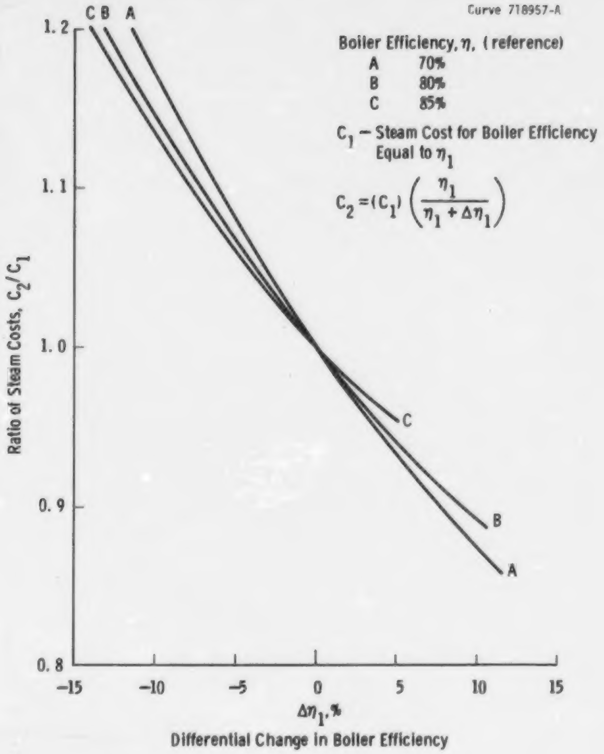

Figure 39 - Steam Costs as a Function of Boiler Efficiency

then these may result in increased capital and annual costs. If such changes affect the combustion efficiency and sulfur capture processes, then the cost impact could be significant.

The effect of the cost of particulate emission control on industrial FBC boilers and conventional boiler plants could be similar for low-sulfur coals. The higher sorbent consumption for eastern high-sulfur coal, however, could result in increased solids loading, and there may be some cost differences between the FBC boiler and conventional boiler plants with wet-scrubbers. The final particulate emission control device - the fabric filter - has not yet been demonstrated in large-scale AFBC operations. At the moderate control level, the final particulate control device efficiency requirements are such that multitube cyclones may be used. The multitube cyclones could be less expensive than the fabric filters but may/or may not be efficient in collecting very fine particles, e.g., smaller than 5 μm. Adequate information concerning

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the particle size distribution in the gas at the primary cyclone exhaust from large-scale AFBC operations is not yet available and, thus, the four reference FBC boiler plants in this study have been designed with fabric filters. It may be possible to optimize the plant design for minimum capital and annual costs.

Impact of Capital Investment on Steam Costs

Significant reductions in total capital investment may be possible for the smaller plants because the ratio of plant auxiliary component cost (material handling systems, etc.) to the FBC boiler cost decreases with an increase in the plant size. The coal handling system costs for the 8.8 and 58.6 MW_t plants are about 20 and 9 percent, respectively, of the total capital investment. If crushed, sized, and dried coal and sorbent are bought, then the coal and sorbent handling system cost could be reduced; but this procedure may be desirable only for the smaller plants. The total capital charges account for approximately 32 to 35 percent of the steam costs, and a 10 percent reduction in the capital investment without incurring additional operating cost could reduce the steam costs by about 4 percent.

Summary of Cost Sensitivity

The effects of the key industrial FBC boiler design and operating variables and the key cost parameters (coal cost and sorbent cost) are summarized in Table 28. The base case value, parametric value, and maximum impact on the boiler efficiency of each parameter are shown. The resulting increase in steam cost for the 8.8 and 58.6 MW_t industrial FBC boiler plants is given. In the case of the fines recycle parameter, a maximum decrease in boiler efficiency of 10 percent is assumed because of a corresponding increase in combustion efficiency. In the velocity, bed depth, and excess air parameters,no change in combustion efficiency was assumed, and boiler efficiency changes are due to changes in stack losses and sorbent calcination energy losses.

Table 28

IMPACT OF STUDY VARIABLES ON STEAM COSTS[a]

Parameter	Base Bed	Parametric Case	$\eta_2-\eta_1$*, %	Percent Increase in Steam Cost	
				8.8 MW_t Plant	58.6 MW_t Plant
Fines Recycle	Yes	No	-10	13.6	13.6
Excess Air, %	20	10	0.31	-0.37	-0.37
		100	-2.09	2.60	2.60
Superficial Velocity, m/s	1.83	1.22	0.08	-0.10	-0.10
		3.05	-1.39	1.69	1.69
Expanded Bed Depth, m	1.22	0.91	-0.64	0.76	0.76
		1.83	0.12	1.06	1.06
Coal Cost, \$/Mg	17	20	0.00	1.80	2.80
Sorbent Cost, \$/Mg(\$/ton)	28 (25)	7 (6)	0.00	-4.70	-7.50

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[a] Bituminous High-S Coal; 1000 μm Average Bed Particle Diameter: High-Reactivity Sorbent; 90% S Removal

Reactivity

*η_1 = boiler efficiency for base case, %

η_2 = boiler efficiency for parametric case, %

In the case of bituminous high-sulfur coal, changing the sorbent cost from \$28/Mg (\$25/ton) to \$7/Mg (\$6/ton) decreases the steam costs by about 5 to 8 percent, depending on the plant size. Similar sorbent cost reductions decrease steam costs by only 0.8 to 1.5 percent for the two low-sulfur coals. For a fixed sorbent cost these results also qualitatively reflect the effect of sorbent consumption, Ca/S molar ratio requirements (sorbent reactivity category), coal sulfur content, and the SO_2 emission control standards on the steam costs. In comparison with some fluidized-bed operating variables, the coal cost, also, has a significant effect on the steam costs.

CONCLUSIONS

As compared with costs for conventional, industrial, coal-fired boiler plants with FGD systems, the capital investment and annual costs of the larger plant capacities for the industrial FBC boiler plants are lower, especially for the western subbituminous coal. The uncertainty, however, in the absolute costs is large enough that the cost comparison can indicate only that the costs are comparable. For a given plant capacity and fluidized-bed operating condition, the annual costs defined per unit of thermal energy output depend strongly on the coal sulfur content and on coal and sorbent costs. For the three reference coals the steam conditions, saturated or superheated, affect the steam costs more than does SO_2 control level. This difference is due to the increased capital investment for generating superheated steam and lower throughputs of superheated steam (kg steam/W_t) than of saturated steam.

Fluidized-bed design and operating variables, such as the carry-over fines recycle option, the number of coal feed points and location, bed temperature, bed depth, superficial velocity, excess air, and sorbent particle size, have a more significant effect on the annual cost per energy output (for e.g., steam cost) than on the total capital investment. These variables influence the combustion efficiency, sorbent calcination energy requirements, stack losses, and auxiliary power consumption, thus affecting boiler as well as plant efficiencies.

The SO_2 emission control standards have little impact on the steam costs (less than 1 percent) for bituminous, low-sulfur (0.9 wt % S) and western subbituminous (0.6 wt % S) coals. Depending on the sorbent activity, however, the steam costs for bituminous high-sulfur coal (3.5 wt % S) at the stringent control level (90% S removal) are 4 to 10 percent higher than at the moderate control level (78.7% S removal, 516 ng SO_2/J). If no sulfur removal were required, a significant reduction in the cost of steam could be realized for the bituminous high-sulfur coal, but only a limited reduction for the low-sulfur coals.

With confidence in the predicted cost trends, we concluded that the three alternative emission levels of SO_2 investigated are constraints not critical to the commercialization of industrial AFBC. The same is true of the particulate control standards investigated and the NO_x standard, although the stringent NO_x standard probably sets a lower limit than more involved NO_x control techniques might later require, with potentially significant cost impact.

We do not expect that changes in the cost premises, such as an increase in the capital recovery factor, will alter the cost trends for industrial AFBC, because the NSPS is seen to have such a small influence on the capital investment. The major conclusions, therefore, will not change with changes in these premises. On the other hand, such changes could alter the absolute cost comparison between AFBC and conventional boilers and shift the advantage from one to the other. Because of the uncertainty in the absolute costs and the sensitivity of the costs to a large number of parameters, we cannot provide quantitative perspective on an absolute cost comparison. Individual plant sites and constraints must be considered for each specific case.

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APPENDIX A

COMMERCIALLY AVAILABLE COAL-FIRED INDUSTRIAL FLUID-BED BOILER DESIGNS

Westinghouse/Zurn Industries[1] and Exxon Research and Engineering Company[2] have carried out conceptual design studies for industrial-sized coal-fired FBC boiler plants. Currently, several boiler manufacturers are participating in demonstration projects and/or offering commerical industrial-sized fluidized-bed steam generation systems. The vendors claim multifuel-firing (coal, oil, and gas) capabilities for the commercially marketed FBC boilers. The reliability and availability of the coal-fired FBC boilers, however, are yet to be demonstrated. Some of the currently active coal-fired FBC boiler demonstration and test units are listed in Table A1. Complete information on all the designs is not available, but some general features of the designs proposed by the major boiler manufacturers in the U. S. are compared in Table A2. The FBC boiler designs proposed differ mainly in the fluidized-bed arrangement and configuration, heat transfer surface layout, and solids feeding technique.

Conventional boilers are designed with superheaters in the convective section, and some FBC boiler manufacturers advocate that the superheating surface be submerged in the bed so as to take advantage of the relatively higher heat transfer coefficient and temperature differential. The arguments raised against in-bed location of superheaters concern localized hot spots, in-bed tube corrosion, and load response capability. The strength of carbon steels and low-alloy steels decreases markedly at temperatures above 427°C (800°F), and the more expensive medium-alloy ferritic and austenitic steels show a rapid decrease in strength above 537°C (1000°F). The materials problem, therefore, dictates careful

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consideration of the localized heat transfer variations and their impact on tube wall temperatures and corrosion/erosion phenomena. In the case of above-bed location, the relatively lower heat transfer coefficient and the smaller driving force (temperature differential) can result in considerably larger superheating surface requirements. The relatively higher dust loadings to the gas leaving the fluidized-bed could increase erosion as well as foul the heat transfer surface located in the convective section. Information from large-scale units is needed to evaluate localized heat transfer variations (especially for segmentally fluidized beds) and corrosion/erosion phenomena to justify either in-bed or above-bed location of superheaters. In the light of the limited data available, however, enough mixing can be expected in the FBC boilers, relative to corrosion/erosion, and localized heat transfer variations may not be a serious problem. Natural as well as forced circulation concepts are being pursued for steam/water tubing layout. Water flow predictions for natural circulation designs are uncertain, and departure from nucleate boiling, steam blanketing, and stratification can occur, resulting in higher tube-wall temperatures that may not be acceptable from a materials point of view.

The FBC boiler designs being offered are generally based on the water-tube heating surface concept. Johnston Boiler Company[4] is offering fluidized-bed boilers with capacities of up to 50,000 lb/hr* steam with water-walled enclosure for the bed and fire-tube heating surface above the bed. Some boiler manufacturers offer tapered bed designs with high velocities at the bottom and relatively lower velocities in the expanded section at the top. This approach can minimize elutriation losses because of the lower velocities at the top but may not provide enough gas residence time for sulfur removal in the lower portions of the bed. Fluidized-bed boilers with more than one bed per module are being

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*Specifications are in British units, as received from vendor.

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Table A1

AFBC COAL-FIRED DEMONSTRATION AND TEST UNITS

(Taken from Reference 3)

Dwg. 2624C42

Developer	Capacity	Location	Sponsor	Status and Comments
Industrial Demonstration Units				
Combustion Engineering	22,700 kg hr (50,000 lb hr) steam	Great Lakes Naval Training Base, IL	U.S. Department of Energy (Cosponsor)	Construction to begin during winter, 1978
Foster-Wheeler; Pope, Evans and Robbins; and Georgetown University	45,400 kg hr (100,000 lb hr) steam	Georgetown University Washington, DC	U.S. Department of Energy(Cosponsor)	Startup scheduled for summer, 1979
Exxon Research and Engineering Co.	3 - 4 MW$_t$ (10-15 × 10^6 Btu/hr)	Linden, NJ or Baytown, TX	U.S. Department of Energy (Cosponsor)	This unit is a process crude oil heater and is currently in the pretesting and design evaluation phase
Battelle-Columbus Laboratories	11,350 kg/hr (25,000 lb/hr) steam	Columbus, OH	U.S. Department of Energy (Cosponsor)	Under design based on SSEUS test unit, see below
Foster-Wheeler; Pope Evans, and Robbins	45,400 kg/hr (100,000 lb/hr)	Wilkes-Barre, PA	U.S. Department of Energy	To be constructed under DOE anthracite culm program
Shamokin Area Industrial Corporation (SAIC); Curtiss-Wright, Dorr-Oliver; Stone and Webster	9,080 kg/hr 20,000 lb/hr) steam	Cellu Products (Paper Company) Shamokin, PA	U.S. Department of Energy	To be constructed under DOE anthracite culm program
FluidDyne Engineering Co.; Deltrak; Nebraska Boiler Company	9,080 - 13,600 kg/hr (20,000-90,000 lb/hr) steam	GTE Sylvania, Towanda, PA	U.S. Department of Energy	Replacement for existing oil-fired oiler, to be constructed under DOE anthracite culm program
Foster-Wheeler; Pope Evans, and Robbins	88 MW$_t$ (220 × 10^6 Btu/hr)	Rivesville, West VA	U.S. Department of Energy	Currently operating
Babcock and Wilcox, Ltd. (England)	12 MW$_t$ (40 × 10^6 Btu/hr)	Renfrew, Scotland	–	Retrofit unit currently operating
Babcock and Wilcox Company (U.S.)	6 MW$_t$ (20 × 10^6 Btu/hr)	Alliance, OH	Electric Power Research Institute	Currently operating
Morgantown Energy Research Center	18 MW$_t$ (60 × 10^6 Btu/hr)	Morgantown, West VA	U.S. Department of Energy	Under design
Babcock and Wilcox, Ltd. (England)	25,250 kg/hr (60,000 lb/hr) steam	Central Ohio Psychiatric Hospital Columbus, OH	Ohio Department of Energy	Startup scheduled for late 1979
To be negotiated	45,400 kg/hr (100,000 lb/hr) steam	Ohio State Penitentary Columbus, OH	Ohio Department of Energy	Retrofit installation currently in planning stage
To be negotiated	160,000 kg/hr (353,000 lb/hr) steam	Columbus and Southern Ohio Electric Company Piqua, OH	Ohio Department of Energy	Utility boiler currently in planning stage
Johnston Boiler Co.	3 MW$_t$ (10 × 10^6 Btu/hr)	Johnston Boiler Co. Ferrysburg, MG	Private	Currently operating
Johnston Boiler Co.	18,200 kg/hr (40,000 lb/hr) steam	Central Soya	Private	Recently sold
Johnston Boiler Co.	9,080 kg/hr (20,000 lb/hr) steam	IBM, Charlotte, NC	Private	Recently sold; designed as oil/gas unit capable of burning coal
Wormser Engineering,	6 MW$_t$ (20 × 10^6 Btu/hr)	Lowell, MA	Private	Currently operating
Mustad and Son (Gjovik, Norway)	25 MW$_t$ (85 × 10^6 Btu/hr)	Varmeverk (Heating Works) Enkoping, Sweden	–	Startup currently scheduled
Coal Processing Consultants (B&W, Ltd.)	36,300 kg/hr (80,000 lb/hr) steam	Prince Edward Island	Private	Startup scheduled for 1982

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Table A1 (Continued)

Dwg. 2624C43

Developer	Capacity	Location	Sponsor	Status and Comments
Industrial Demonstration Units				
Energy Equipment	13,600 kg/hr (30,000 lb/hr) steam	Cadbury SCHWPS Boarnville, UK	Private	Currently operational
Ruhrkohle	35 MW_t (105×10^6 Btu/hr)	Düsseldorf-Flingern West Germany	Private	Startup scheduled for early 1979
Wesertal GMBH	125 MW_t (375×10^6 Btu/hr)	Hameln, West Germany	Private	Currently operational
Mitchel Engineering	36,300 kg/hr (80,000 lb/hr) steam	Don River, UK	British Steel	Currently operational
Pilot-Scale Test Units				
Combustion Engineering	1 MW_t (3×10^5 Btu/hr)	Windsor, CT	U.S. Department of Energy	Currently operating
Energy Resources Company	1.8 MW_t (6×10^6 Btu/hr)	Cambridge, MA	Private	Currently operating
Pope, Evans and Robbins	1.5 MW_t (5×10^5 Btu/hr)	Alexandria, VA	U.S. Department of Energy	Currently operating
Stone Platt Fluidfire, Ltd.	0.3 MW_t (1×10^6 Btu/hr)	Virginia Polytechnic Institute; Blackburg, VA	–	Startup currently scheduled
Stal-Laval Turbine Company (Finspaug, Sweden)	1.5 MW_t (4.5×10^6 Btu/hr)	District Heating Plant Orebro, Sweden	–	Currently operating
FluiDyne Engineering	55,700 kg/hr (12,600 lb/hr) hot air output	Minneapolis, MN		
EPA Sampling and Analysis	0.3 MW_t (1×10^6 Btu/hr)	Research Triangle Park, NC	U.S. Environmental Protection Agency	Currently operating
Babcock and Wilcox Company (U.S.)	1.5 MW_t (5×10^6 Btu/hr)	Alliance, OH	Electric Power Research Institute	Currently operating
Battelle-Columbus Laboratories (SSEUS)	1.5 MW_t (5×10^6 Btu/hr)	Columbus, OH	U.S. Department of Energy	Currently operating

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Table A2

INDUSTRIAL-SIZED FLUID-BED BOILER DESIGNS[a]

Boiler Manufacturer/Developer and Sponsor	Foster Wheeler/Pope, Evans and Robbins, and U.S. D.O.E.	Combustion Engineering/ U.S. D.O.E.	Babcock & Wilcox U.S.A.	Johnston Boiler Co., U.S.A.	FluiDyne Engineering Co., U.S.A.
Plant Details					
Steam capacity	100,000 lb/hr steam	50,000 lb/hr steam	200,000 lb/hr steam	2,500 to 50,000 lb/hr steam	Air heater
Location	Georgetown University, Washington, DC	Great Lakes Naval Training Base, IL	Conceptual	Commercially available	Commercially available
Steam conditions	625 psig, saturated	365 psig, 500°F	150 psig, saturated	Up to 300 psig	
Number of beds module	2	8 zones	5	Tapered bed with expanded section at the top	
Arrangement of beds	Dual bed with partitioning wall	Air plenum divided into eight zones	Segmented bed	Segmented bed with water tube bed dividers	
Boiler dimensions	19 ft 4 in x 11 ft (2 segments)	13 ft wide, 19 ft long and 18 ft deep	NA[b]	12.75 ft wide, 31.7 ft long and 16.1 ft deep (50,000 lb/hr boiler)	
Expanded bed depth, ft	4.5	3	4	2.67	3.5 - 4.0
Temperature, °F	1594	1550	1550	1500-1650	1325 - 1465
Superficial velocity, ft/sec	8	7	8	6 to 12 at the bottom and 3 to 6 at the top	2.0 - 4.2
Freeboard height, ft	8				
Excess air, %	10	20	21	25	30 - 130
Ca/S molar ratio[c]	3	4	4	-	3.6
Flue gas temperature at stack inlet, °F	400	NA	300	305	NA
Steam water circuitry	Natural circulation (water-tube)	Natural circulation (water-tube)	Forced circulation (water-tube)	Water cooled combustion Chamber; water-walled enclosure with fire tube Heating surface above bed	

Table A2 (Continued)

Boiler Manufacturer/Developer and Sponsor	Foster Wheeler/Pope Evans and Robbins, and U.S. D.O.E.	Combustion Engineering/ U.S. D.O.E.	Babcock & Wilcox U.S.A.	Johnston Boiler Co., U.S.A.	Fluidyne Engineering Co., U.S.A.
Plant Details					
In-bed heat transfer tubes	Inclined	Inclined A-type	Horizontal tubes		**Horizontal tubes**
Control of carbon utilization	Recycle of fines	Recycle of fines	Recycle of fines	Recycle of fines	**Recycle of fines**
Coal feeding technique	Overbed; Stoker (2) side-wall mount	In-bed feeding, one feed nozzle zone	In-bed feeding; feed point per 9 sq. ft	Over-bed coal fired	
Sorbent feeding technique	Overbed; 4-in dia feed pipes (2)	Premixed with coal	In-bed	Premixed with coal	In-bed
Bed overflow	One exit point bed; gravity drain	One exit point per zone; gravity drain	Gravity drain	Gravity drain	**Proprietary bed drain technique**
Coal type[c]	Bituminous	Illinois No. 6	NA	NA	
Heating value	12,750 Btu/lb	10,430 Btu/lb	11,000 Btu/lb	NA	
Ash	7.97	9.2	10.0	NA	
Moisture, %	5.00	17.00	NA	NA	
Sulfur, %	3.29	3.50	3.0	NA	
Feed size	1 in x 1/4-in	>1.5 in	>7 1/4 in	Between 0.25 to 1.25 in	
Sorbent type[c]	Greer/Grove	NA	NA	NA	NA
Sorbent particle size	1/8 in by 16 mesh	NA	>3/8 in	8 x 16 mesh	1/4 in
Coal flow rate, lb/hr	9565	6710	NA	NA	
Sorbent flow rate, lb/hr	3133	3260	6500	NA	
Recycle rate, lb/hr	7500	5640	NA	NA	
Thermal efficiency, %	83.51	81.8	84.0	81	
Enrironmental Performance					
Percent sulfur capture	90	90	85+	NA	
Final Particulate Control Device	Baghouse	Baghouse	Baghouse	Baghouse	**Baghouse**

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[a] Specifications in British units are as received from vendor.

[b] NA - Adequate information not available.

[c] Vendor design basis.

designed to handle the various load control requirements. Some FBC designers use a partitioning wall for separating the bed, and others divide the air plenum so as to obtain segmental fluidization.

Three European companies, Babcock & Wilcox Ltd. (England), Stone-Platt FluidFire Ltd., and Mustad and Son (Norway) are also actively involved in the design, development, and marketing of FBC boilers. Stone-Platt has currently designed 10 AFBC units in the size of 0.1 to 7 MW_e and has built three prototypes of 0.1 MW_e (hot water generation) and a 5,000 lb/hr steam prototype plant. The bed is concave, with the heat exchange surface located in the shallower portions of the bed. The design is based on coal and limestone feed, using a screw-feed system with one feed point per bed, and requires the use of 250 to 750 μm sorbent particles. The feed point location is such that coal is introduced just below the top surface of the bed at the center of the unit. Babcock & Wilcox Ltd. is offering AFBC designs in the range of 100,000 to 600,000 lb/hr steam and has several years experience with various FBC units, including those at Renfrew, Scotland (40,000 lb/hr steam). It is also participating in demonstration programs in the U. S. (Table A1). The bed in its design is compartmentalized and is fed with premixed coal (<0.63 cm [-1/4 in.]) and sorbent (0.32 to 0.63 cm [1/8 to 1/4-in.]) using multiple pneumatic feed points. Mustad offers fluidized-bed boilers in the range of 4.5 to 2.2 MW thermal input, and its design utilizes preheated combustion air with provision for secondary combustion air above the bed. Problems associated with solids feeding, heat transfer tube corrosion and/or erosion, etc. have occurred with these designs, and further work is being performed to design reliable coal-fired AFBC systems.

Both in-bed and above-bed feeding techniques are being considered in the fluidized-bed boiler designs offered in the U.S. In the light of stable operation experienced at the Rivesville plant, with some of the feed nozzles remaining plugged, in-bed feeding systems with greater spacing (one feed point per 18 sq. ft compared to 9 sq. ft) have been suggested. Uniform distribution of coal feed, however, is still a technical design uncertainty and can affect the sulfur release and capture phenomena in the bed. An over-bed feeding technique has been proposed to overcome some of the problems associated with in-bed feeding, like feed-nozzle plugging, coal-feed distribution, etc. This technique can result in combustion, however, and, hence, sulfur release in the freeboard, and can affect sorbent utilization and combustion efficiency. Recycle of fines to the main bed is being considered for improved carbon utilization in industrial-sized boilers and seems to be preferable to a separate carbon burnup cell for economic and technical reasons.

The operating conditions being considered for industrial FBC boilers include superficial velocities of 1.22 to 2.43 m/s (4 to 8 ft/sec) with excess air in the range of 10 to 20 percent and expanded bed depths between 0.61 and 1.37 m (2 and 4.5 ft). With the gas residence times and sorbent particle sizes being considered in the boiler designs compared in Table A2 higher Ca/S ratios than those assumed may be required.

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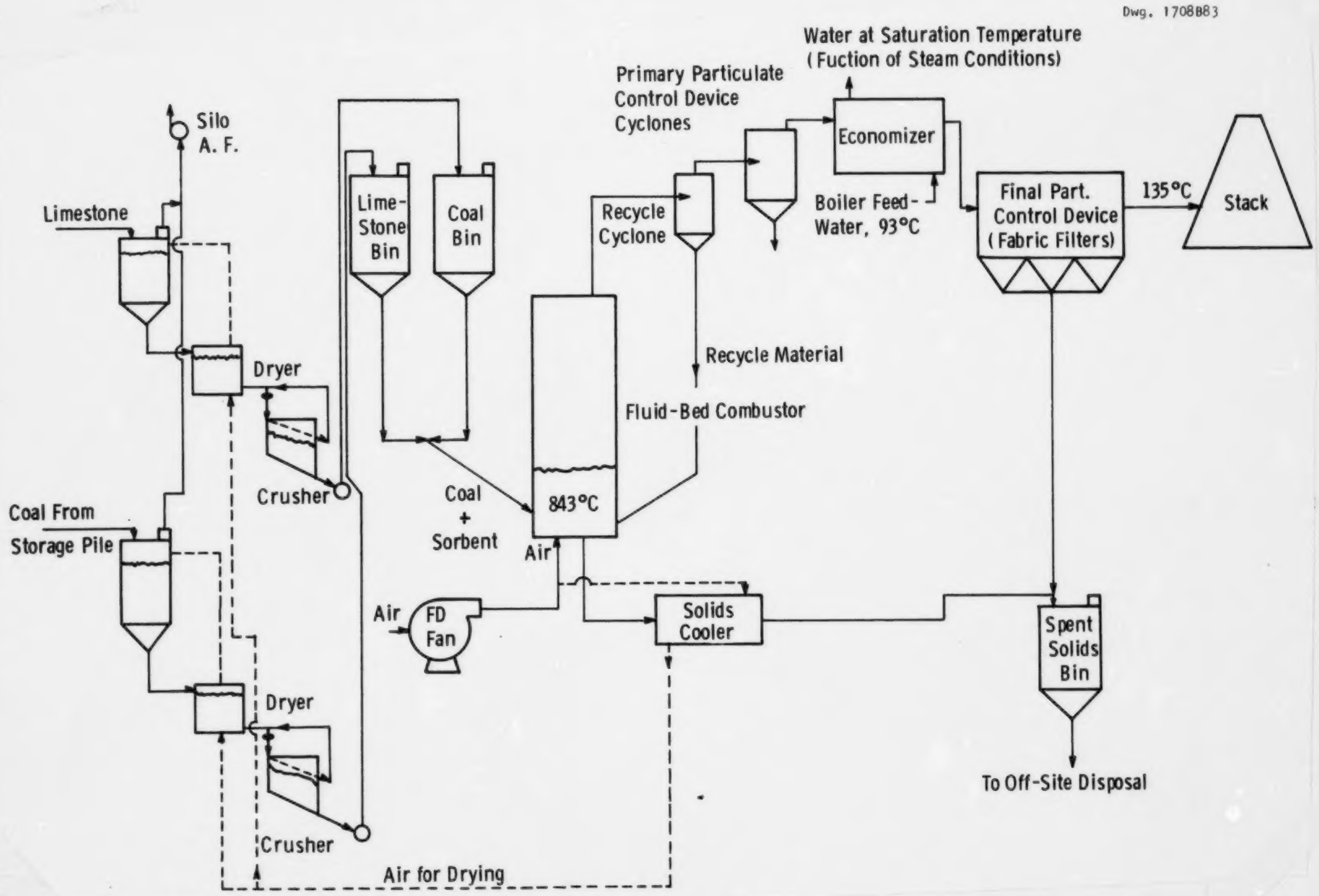
APPENDIX B

INDUSTRIAL FLUIDIZED-BED BOILER DESIGN DESCRIPTION

A conceptual design of an industrial fluidized-bed combustion boiler system has been prepared. The plant components are shown schematically in Figure B1. The fluidized-bed combustor is of dual bed design with air supplied by forced draft fans split into parallel streams. Airflow to the individual beds is controlled with dampers and air measurement devices in the air ducts to the windboxes. Coal and sorbent, premixed, are fed into the water-walled enclosed beds, with submerged heat transfer surface for maintaining constant bed temperature. The combustion products leaving the fluidized bed pass through the convective section before entering the mechanical cyclone collectors (primary particulate control device), and the solids collected are reinjected into the combustor bed for improved carbon and sorbent utilization. Heat is recovered from the bed overflow solids in an air heater and used for drying the solid feed streams to the bed. An economizer is located between the primary and final particulate control system for boiler feedwater preheat. The flue gas leaves the final particulate control device at a temperature of 135°C (275°F) to the stack. The bed overflow material and the fines collected by the final particulate control system (a fabric filter) are stored for final off-site disposal.

FLUIDIZED-BED STEAM GENERATOR

The atmospheric-pressure fluidized-bed combustor consists of two beds enclosed within a water-wall shell and without a partitioning wall between the beds. The air plenum is divided into two (or more) zones to achieve segmental fluidization by local bed slumping. Water walls are designed for natural circulation with 5.08 cm (2-in) diameter tubes on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Recycle Material
Primary Particulate
Control Device
Cyclones
Solids
Cooler
Water at Saturation Temperature
(Fuction of Steam Conditions)
Economizer
Boiler Feed-
Water, 93°C
Final Part.
Control Device
(Fabric Filters)
135°C
Stack
Spent
Solids
Bin
To Off-Site Disposal

Figure B1 - Industrial FBC Boiler Plant - Process Flow Schematic

7.6 cm (3 in) centers. The submerged horizontal heat transfer surface is designed for forced circulation of water and consists of 2.5 cm horizontal tubes with 7.5 cm by 15.0 cm diamond spacing. A tube-free volume is provided between the submerged surface and the grid (0.3 m) and between the top row of the submerged surface and the splash screen in the freeboard (0.22 m). The particle splash screen incorporates three rows of 2.5 cm tubes on 7.5 cm by 7.6 cm (3 in) diamond spacing, permitting operation with lower freeboard heights. In-bed feeders (pneumatic injectors) are provided for feeding premixed coal and sorbent. Provision is made for reinjecting fines into the combustor bed.

In the convective pass 5.1 cm tubes are arranged on 11.0 cm spacing to recover heat from the flue gas leaving the fluidized bed. The steam/water circuitry is designed for natural circulation, and the heat transfer surface is located in an idle pass within the boiler module, before the primary particulate control device.

The base operating conditions for the FBC boiler are: temperature, 843°C; excess air, 20 percent; superficial gas velocity, 1.83 m/s; and expanded bed depth, 1.22 m. The four reference plants have been designed with one or more boiler modules so that they can be operated with superficial gas velocities of 1.22 to 2.44 m/s, excess air levels of 10 to 100 percent, and expanded bed depths in the range of 0.91 to 1.83 m. The maximum boiler dimensions are constrained by shipping limitations. The 8.8 and 22 MW_t plants have been designed with one boiler module, while the 44 and 58.6 MW_t plants contain two and three modules, respectively. The 58.6 MW_t plant, however, could be designed with two boiler modules of 25.3 MW_t capacity each, as shown in Figure B2. The shop-fabricated boiler module dimensions could be such that the superficial velocity could not be lowered to 1.22 m/s without changing other operating variables. Velocities higher than those of the base case can result in shortened gas residence time in the bed and could increase sorbent consumption and elutriation of bed material.

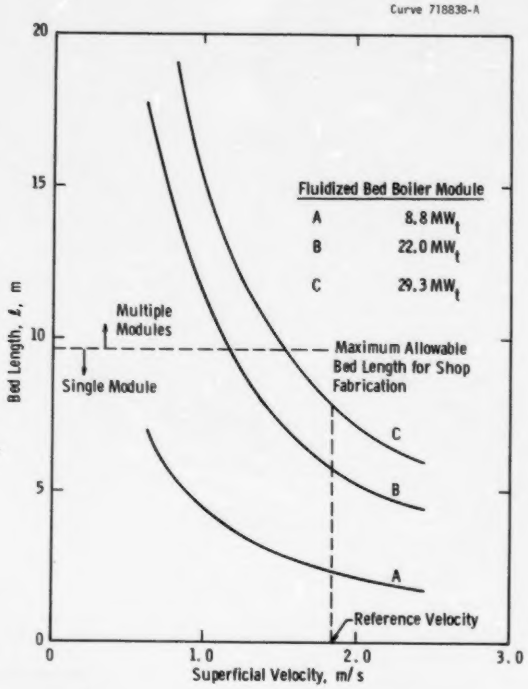

Figure B2 - Effect of Gas Velocity on Number of Boiler Modules

Similarly, operating the FBC boiler with relatively deeper beds (more than 1.22 m) may be undesirable, because it could increase the pressure drop and elutriation losses.

COAL AND SORBENT HANDLING SYSTEMS

Coal and sorbent handling systems are independent of each other and include receiving hoppers, silos, feed elevators, sizing and drying systems, screens, surge bins, bin vibrators, and necessary valves and piping arrangements. Coal and limestone are received from storage and crushed to <0.63 cm (1/4 in) and <0.32 cm (1/8 in) size, respectively. The crushed materials are transferred to their respective feeding hoppers and then to the moving-belt gravimetric feeders with adjustable belt speed for feed rate control. Coal and limestone rates are independently verified, then mixed and fed to the boiler.

CAUTION: Mark corrections in Sky Blue pencil ONLY.

AIR AND FLUE GAS HANDLING SYSTEM

A combination of forced draft (FD) and induced draft (ID) fans, or a single forced draft fan, could be used to meet the gas flow and pressure drop requirements. The draft losses for the base fluidized-bed design and operating conditions could be of the order of 17.2 kPa (0.17 atm ≈68 in w.g.). A balanced draft system may be preferable to avoid gas leakage and to operate the FBC boiler close to atmospheric pressure. The balanced draft system (FD and ID fans) could be relatively more expensive than the forced draft fan system. When the operational and maintenance flexibilities of the two systems are compared, however, the impact of the balanced draft system on the overall plant investment could be small.

SPENT SOLIDS HANDLING SYSTEM

The fluidized-bed combustor is provided with one exit point per bed, and teh bed material drainage rate is controlled using starwheel, high-temperature valves into a rotating drum-type solids cooler. Heated air from the solids cooler is used for drying the solids feed streams to the boiler. The cooled solids are then conveyed with an air-lock transport system to a storage silo for final off-site disposal.

PARTICULATE CONTROL SYSTEM

Mechanical cyclone collectors and fabric filters are used for controlling particulate emissions. The particulate (fly ash) handling system beneath the collector and fabric-filter outlets is of conventional design and consists of surge bins and high-temperature air-lock valves discharging into a pneumatic transport system. The conveying lines can be used for transporting the fines back to the combustor bed for reinjection and/or to the storage silo for final off-site disposal with the bed overflow material.

The effect of controlling load changes by varying fluidized-bed operating conditions (temperature, velocity [excess air], bed depth) can influence sorbent requirements (due to changes in gas residence time)

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for a given SO_2 emission control level. Changes in the feed rates and operating conditions can also affect the particulate loadings to the flue gas (elutriation, attrition, and recycle rates) and the particle size distribution in the bed.

OPERATING TECHNIQUES

Start-up

The start-up from a cold condition is accomplished by auxiliary fuel firing and segmental fluidization of the bed. First, the bed in the static mode can be preheated by allowing combustion of an auxiliary gaseous fuel into/or above the start-up zone. At around 550°C (1022°F), start-up oil can be injected while the start-up zone is minimally fluidized. Coal is then admitted and ignited, and the auxiliary fuel input is gradually stopped. Start-up of other portions of the bed, as well as of the adjacent bed, can rely on solid recirculation from the operating zone. Airflow to the start-up zone is controlled with dampers in the windboxes.

Load Control

Turning-off one of the beds can provide a turndown ratio of at least 2:1 for the entire boiler. A change of bed temperature fuel input rate, superficial velocity, and excess air, either individually or in combination, can provide up to 30 to 50 percent load reduction capability. Hence, an overall turndown of 4:1 can be accomplished.

The effect of controlling load changes by varying fluidized-bed operating conditions (temperature, velocity [excess air], bed depth) can influence sorbent requirements (due to changes in gas residence time) for a given SO_2 emission control level. Changes in the feed rates and operating conditions can also affect the particulate loadings to the flue gas (elutriation, attrition, and recycle rates) and the particle size distribution in the bed.

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APPENDIX C

SULFUR REMOVAL PROJECTIONS

The projections of calcium-to-sulfur molar feed ratio required to achieve any selected degree of desulfurization in a fluidized-bed combustor are derived by using a simplified model for fluidized-bed desulfurization, with kinetics rate constants developed from laboratory thermogravimetric (TG) data. The TG data base, the model, and its agreement with fluidized-bed data have been discussed previously.

THE WESTINGHOUSE MODEL

The model used here has been described previously (Appendix E of Reference C1). The following assumptions are made in deriving the model:

1. Release of sulfur from coal as sulfur dioxide (SO_2) occurs with equal probability at any place in the bed in AFBC.
2. Sulfur dioxide released at a distance, $\bar{h}$, below the upper bed surface passes in plug-flow through a homogeneous bed of sorbent of height h. If a mean reaction rate, k, can be assigned to the absorption of gas in the bed, the fraction of the sulfur retained is $1 - e^{-kt}$, where t is the gas residence time in the bed of height h. Integration for all points $\bar{h}$ along the bed height yields an expression for the fractional sulfur removal. R, $R = 1 - \frac{1}{kz}(1 - e^{-kz})$, where z is the residence time of gas entering the bed from below.
3. The mean rate constant for gas absorption in the bed, k, is obtained at any fixed mean utilization of the sorbent from thermogravimetric data reduced to unit gas concentration. The

rate constant for any given sorbent type and chemical form will depend on the mean sorbent utilization, sorbent particle size, bed temperature, bed pressure, and gas composition in the bed.

The rate constant for limestone sulfation required for various sulfur removal efficiencies is projected in Figure C1 as a function of gas residence time. In practice, the rate constant could be varied by changing, for example, the sorbent particle size or bed temperature. Alternatively, bed residence time could be adjusted by varying gas velocity or bed depth.

Operating conditions that would disrupt the uniform sulfur generation pattern are: the location of coal feed points and their number, the coal particle size, and the extent of vertical mixing in the operating fluidizing mode. Fine particles of coal fed near the base of a tall bed at low fluidization velocity would, thus, probably cause disproportionate release of SO_2 near the base of the bed and improve sulfur removal, relative to the prediction of a uniform sulfur generation model.

TG DATA

The use of TG data to model the fluidized-bed sulfation depends on several assumptions, which are listed below:

1. The rate-limiting process is governed by diffusion within the sorbent itself.
2. The residence time of the sorbent in the bed does not affect its reactivity (although the degree of sulfation does affect reactivity).
3. The fluidized-bed atmosphere surrounding the limestone particles is oxidizing with respect to the $CaS/CaSO_4$ transition.

The implication of the first assumption is that while mass transfer (e.g., of SO_2 from the gas phase to the sorbent particle) in the fluidized bed may be much greater than in the TG apparatus, mass transfer dominates the actual reaction rate over a small extent of reaction.

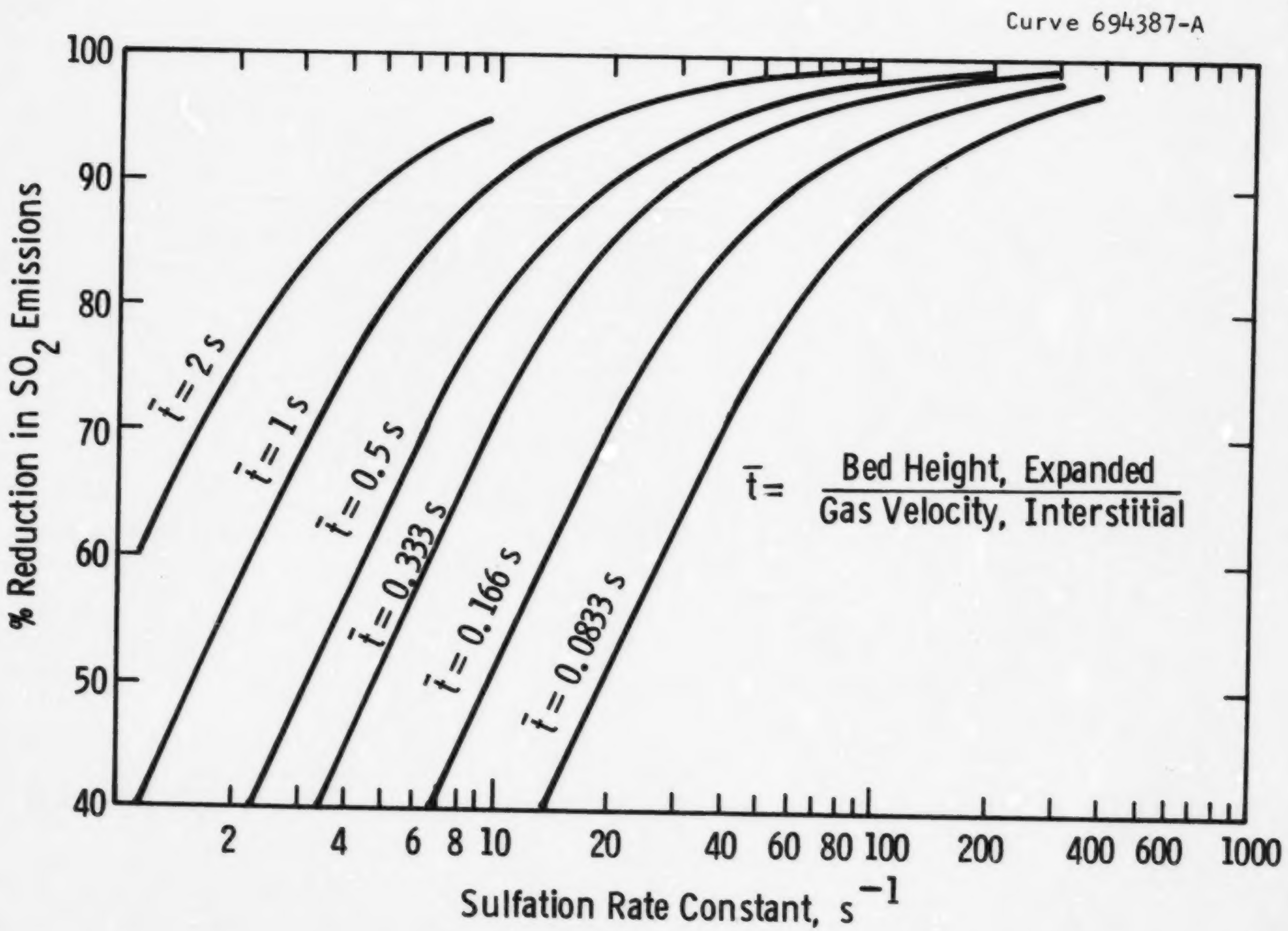

Figure C1 - Required Reaction Rate as a Function of Sulfur Removal Efficiency and Gas Residence Time ($\bar{t}$)

CAUTION: Mark corrections in Sky-Blue pencil ONLY.

Once the calcium utilization has reached about 10 percent, the rate is usually dominated instead by inter- and intragranular diffusion of the solid. The same internal pore structure must be provided in the sorbent during the TG test as would prevail during a fluidized-bed combustor operation. Thus, the calcination conditions under which the sorbent pore structure is generated must be carefully controlled before the TG sulfation experiment if the TG data are to provide a valid simulation of the effect of sorbent utilization on reaction rate.

Concerning the second assumption above, the mean residence time of the sorbent may be much longer in a fluidized bed (up to 50 hours) than in a TG experiment (∿2 hours), unless the latter is carried out under conditions of low-SO_2 concentration (0.05%). The sorbent pore structure may be modified by this long exposure to temperature, and, unfortunately, this may either enhance or retard the sulfation rate, depending on the particular sorbent. This effect is not accounted for in the TG-based kinetic model.

It is often postulated that the existence of local reducing areas in the bed may cause sulfur capture as CaS, which is subsequently oxidized to sulfate. The assumption that this mechanism is unimportant in determining the overall rate of reaction is justified by the fact that oxidation of CaS in limestone is limited in the same way that utilization of CaO is limited in sulfation. Oxidized sulfided limestone contains an inner layer of sulfided stone that has not been oxidized. Since CaS is not found in the product stone from FBC, it is clear that there is no significant additional reaction rate component from sulfidation of the sorbent.

The TG data, which yield rate as a function of sulfur loading or sorbent utilization, vary greatly according to the sorbent properties and combustor operating conditions. The variables that affect the reaction rate and can be studied on the TGA are:

1. Type of sorbent

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2. Form of sorbent (e.g., limestone, lime, hydrated lime)
3. Particle size of sorbent
4. Calcination conditions (in particular the temperature and partial pressure of CO_2 during calcination)
5. Residence time (time at temperature)
6. System pressure
7. System temperature
8. Percent excess air in the system
9. SO_2 partial pressure and other gas composition variables.

In making a projection of the effect of operating conditions on the calcium-to-sulfur molar feed ratio, it is advisable to use TG data obtained under experimental conditions that correspond closely to those that will be encountered in the fluidized bed.

For any set of design input:

- Bed height
- Gas velocity
- Bed voidage
- Bed emulsion and bubble phase volume
- Coal:
  - Ash content (heating value)
  - Sulfur content

the rate data can be used in the model to give calcium-to-sulfur molar feed ratio projections. A data base of over 300 atmospheric-pressure TG runs is available at Westinghouse over the range of conditions listed in Table C1.

BASE OPERATING CONDITION PROJECTIONS

Sorbent feed projections were obtained for the base operating conditions (Table C2) using TG runs 231, 241, P17 and 75-281. The rate data were fitted with polynomial equations for use in the computer generated projections.

Table C1

RANGE OF ATMOSPHERIC TG SULFATIONS

Sorbent(s)	Size U.S. Mesh	% Excess Air	Temp., °C	Comment
Ames, Brownwood, 1359, Greer, Carbon 1337, Western, Mississippi Bellefonte	16/18 5-100/200	20	815	Varied calcination
Greer	16/18	20	780-950	Varied calcination
Ames	35/40	20	750-940	
Ames, Brownwood Greer, Carbon, 1359	35/40	20	900	Calcined at 900°C in 60% CO_2
Ames, 1359, Greer, Carbon	5/6-35/40	20	850	Scattered particle sizes
1359 (2263)*	16/18	20	815	
Carbon	-325	20	800-950	
Tymochtee	16/18	2-16% O_2	815	
Canaan, Kaiser	100/200	20	815	Calcined at 900°C in 60% CO_2

*Residence time varied (0.1-0.5% SO_2 in sulfating atmosphere).

Table C2

FBC PLANT OPERATING CONDITIONS

	AFBC	
Pressure, kPa	101	
Bed Temperature, °C	840	
Excess Air, %	20	
Sorbent Type	Limestone	
Sorbent Diameter, μm	1000	500
Bed Velocity, m/s	1.83	
Bed Depth, m	1.22	
Gas Residence time, s	0.67	

The average rate constant for SO_2 sorption in the bed, K, that is needed to maintain a given level of sulfur retention, R, was calculated at a defined gas residence time ($\tau = \frac{\text{expanded bed height}}{\text{interstitial gas velocity}}$) using the equation:

$$R = 1 - \frac{1}{K\tau}\left(1 - e^{-K\tau}\right)$$ fraction of released sulfur captured in the combustor bed.

K is related to the first-order rate constant for the reaction $CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$, K', by:

$$K = K' \frac{(1 - \gamma)\ \varepsilon\ (1 - F)}{\gamma + (1 - \gamma)\ \varepsilon},$$

where

$K = f(\alpha)$

γ = volume fraction of bed bubbles

ε = bed voidage in emulsion phase

F = fraction of emulsion volume occupied by inerts

U = interstitial gas velocity $= \dfrac{U_s}{[(1-\gamma)\ \varepsilon + \gamma]\ (1 - \bar{f}_H)}$

U_s = superficial gas velocity

$\bar{f}_H$ = fraction of bed volume occupied by heat transfer surface

$$\gamma = \frac{\varepsilon_{mf}}{1-\varepsilon_{mf}} \left[\frac{1}{C_D\varepsilon_{mf} + (1-\varepsilon_{mf})\ \dfrac{U_{mf}}{U_s}}\right] - 1$$

C_D = constant

ε_{mf} = bed voidage at minimum fluidization

U_{mf} = minimum fluidization velocity

h = expanded bed height

$\tau = h/\pi$.

The sorbent utilization, α at which the reaction rate constant, K' applies, is determined from TG data. A mass balance on the TG system gives the molar rate of reaction, $\frac{d\alpha}{dt}$, in terms of the reaction rate constant as:

$$\frac{d\alpha}{dt} = \frac{K'C\ \varepsilon_{TG}}{p\ (1 - \varepsilon_{TG})} = 2.0185 \times 10^{-6}\ (K'),$$

where

$\varepsilon_{TG} = 0.5$
P = solids density, mole Ca/cc = 0.0271 (for limestones)
C = mole SO_2/cc in TG reaction gas = 5.47×10^{-8} (atmospheric pressure, 0.5% SO_2).

Polynomial equations representing TG rate data $\alpha = f\ (\frac{d\alpha}{dt})$ are used to calculate the sorbent utilization. The required calcium-to-sulfur molar feed ratio, C, is then defined by:

$$C = \frac{\eta R}{\alpha},$$

where η is the fraction of the sulfur fed to the FBC unit which is released from the coal.

For the Ca/S molar ratio projections in the attached tables, the following have been assumed:

$\varepsilon_{mf} = 0.5$
U_{mf} = 0.06 m/s (0.19 ft/sec) for 500 m particles
η = 1.0 (for recycle of fines)
$C_D = 1.3$
U_s = 1.83 m/s (6 ft/sec)
h = 1.22 m (4 ft)
F = 0
ε = 0.5

$\bar{f}_H = 0.05625$ (a function of bed depth and velocity)

$\gamma = 0.502$

$U = 2.59$ m/s (8.466 ft/sec)

$\tau = 0.4725$

$K' = 3.016\ K = 3.016\ (K\tau)(\frac{1}{\tau}) = 6.383\ (K\tau)$ where K' is in $(\text{sec})^{-1}$.

For a sulfur removal efficiency of 90 percent.

$K\tau = 9.0$

$K' = 6.383\ (K\tau) = 57.447\ (\text{sec})^{-1}$

$\frac{d\alpha}{dt} = 2.0185 \times 10^{-6}\ (K') = 1.1596 \times 10^{-4}\ (\text{sec})^{-1} = 6.957 \times 10^{-3}\ (\text{min})^{-1}$.

For high-reactivity sorbent, Western 90 percent CaL Stone and 1000 μm particles:

$$\alpha = f\left(\frac{d\alpha}{dt}\right)_{\frac{d\alpha}{dt} = 6.957 \times 10^{-3}} = 0.24172$$

α for 500 μm particles = α for 1000 μm particles + 0.1032 [ln (1000/5000)]

$$(\alpha)_{500\ \mu\text{m particles}} = (\alpha)_{1000\ \mu\text{m particles}} + 0.1032\ [\ln\ (1000/500)]$$

$$(\alpha)_{500\ \mu\text{m particles}} = 0.24172 + 0.07153 = 0.31325$$

$$C = \frac{\eta R}{\alpha} = \frac{(1)\ 0.9}{0.31325} = 2.873.$$

VALIDITY OF THE ASSESSMENT PROJECTIONS

A computerized file of fluidized-bed combustor data from PER, the NCB, ANL, Exxon, and B&W was searched for combustor runs carried out at operating conditions near the base case conditions for the technical/economic assessment. The limited bench-scale and pilot-plant data

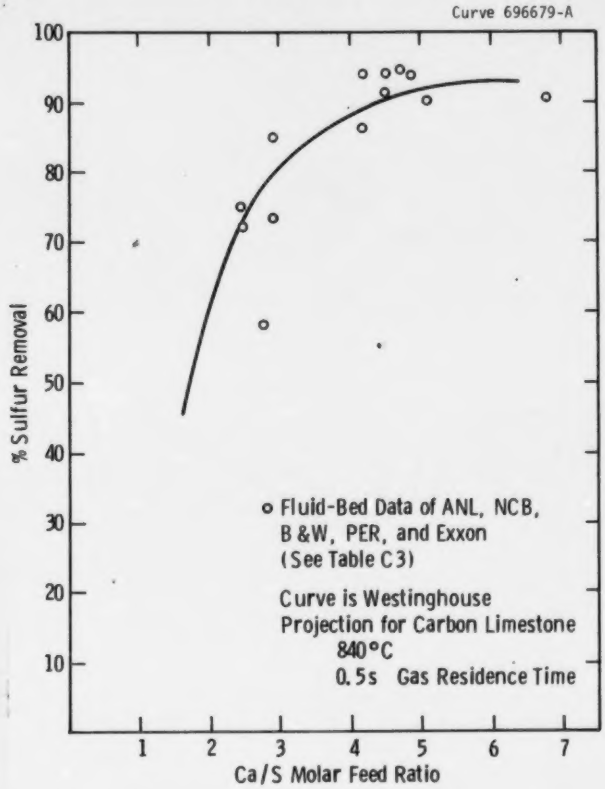

Figure C2 - Fluid-Bed Combustor Data at Conditions near the 1000 μm AFBC Base-Case Projection Conditions

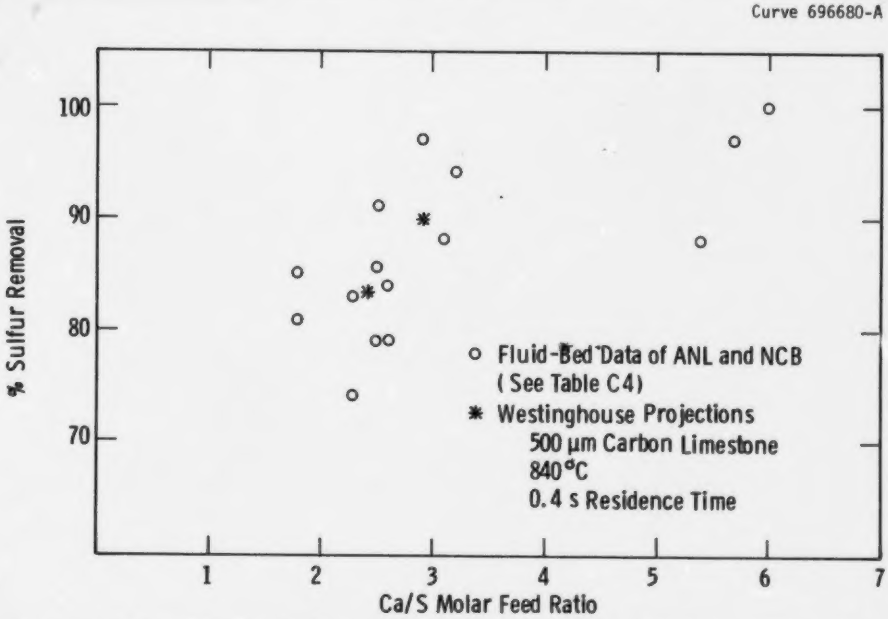

Figure C3 - Fluid-Bed Combustor Data at Conditions near the 500 μm AFBC Base-Case Projection Conditions

Table C3

FLUID-BED COMBUSTOR DATA AT CONDITIONS NEAR THE 1000 μm AFBC BASE CASE PROJECTIONS

Study	Sorbent	Particle Size, μm	Bed Temperature °C	Gas Residence Time, s[a]	Ca/S Molar Ratio	% SO_2 Removal
Continuous-Feed Unit Tests:						
ANL(C2)	1360	1000-1400[b]	871	0.42	2.5	72
ANL(C2)	1360	1000-1400[b]	871	0.42	4.2	86
ANL(C2)	1359	1400[c]	871	0.67	1.6	18
NCB(C3)	18	<3175[b]	777	0.38	2.9	73
B&W(C4)	Carbon	<6350[b]	819	0.19	2.8	58
B&W(C5)	Carbon	1082[d]	854	0.45	4.5	92
B&W(C5)	Carbon	1151[d]	871	0.52	4.6	95
B&W(C5)	Carbon	997[d]	863	0.46	4.8	93
B&W(C5)	Carbon	1206[d]	865	0.38	4.5	93
B&W(C5)	Carbon	1118[d]	867	0.45	4.2	94
B&W(C5)	Carbon	901[d]	839	0.52	2.4	76
B&W(C5)	Carbon	938[d]	844	0.54	2.8	85
Batch Unit Tests:						
PER(C5)	1359	840-2381[b]	860	0.14	16.0	93
Exxon(C7)	1359	930[c]	871	0.25	6.8	90
Exxon(C7)	1359	930[c]	871	0.25	5.4	90
Exxon(C7)	1359	930[c]	982	0.25	5.1	90
Exxon(C7)	1359	930[c]	927	0.28	5.9	90

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[a] $\text{Gas Residence Time} = \dfrac{\text{Bed height expanded}}{\text{Superficial Velocity}}$.

[b] Size range of feed sorbent.

[c] Mass average diameter of feed sorbent.

[d] Surface mean diameter of bed material.

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Table C4

FLUID-BED COMBUSTOR DATA AT CONDITIONS NEAR THE 500 µm AFBC BASE CASE PROJECTIONS

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Study	Sorbent	Particle Size, µm[a]	Bed Temperature, °C	Gas Residence Time, s[b]	Ca/S Molar Ratio	% SO_2 Removal
ANL(C8)	1360	630	871	0.7	2.3	74
ANL(C8)	1359	490	815	0.7	2.5	91
ANL(C8)	1359	490	843	0.7	2.5	91
ANL(C8)	1359	490	871	0.7	2.5	86
ANL(C8)	1359	615	899	0.7	2.6	79
ANL(C8)	1359	615	899	0.7	2.3	83
ANL(C8)	1359	615	899	0.7	2.5	79
ANL(C9)	1359	609	843	0.63	11.9	90
ANL(C9)	1359	609	843	0.66	5.4	88
NCB(C3)	UK	537	799	0.7	3.2	94
NCB(C3)	UK	537	799	0.7	1.8	81
NCB(C3)	UK	537	799	0.7	2.9	97
NCB(C3)	UK	537	799	1.0	1.8	85
NCB(C3)	18	453	848	0.7	5.7	97
NCB(C3)	18	453	848	0.7	6.0	100
NCB(C3)	UK	537	799	1.0	2.6	84
NCB(C3)	UK	537	799	1.0	3.1	88

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9

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[a] Mass average diameter of feed sorbent.

[b] Gas Residence Time = $\frac{\text{Bed height expanded}}{\text{Superficial velocity}}$.

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available at the base case conditions indicate that the TG projections are representative of the fluidized-bed data, although a direct comparison using the three representative limestones from this assessment cannot be made.

Atmospheric combustor data at operating conditions near the base case are given in Tables C3 and C4. Note that the sorbent particle size from most runs is reported only for the feed sorbent and is usually a mass average diameter rather than a surface mean. Figure C2 compares the calcium-to-sulfur feed requirements for desulfurization from Table C3 with the TG projections for 1000 μm particles. The 500 μm data combustor runs are plotted in Figure C3 with TG projections. The scatter of combustor data is partially due to the range of operating conditions plotted, i.e., 799 to 899°C temperature and varied sorbent types. No runs were performed with recycle of cyclone fines.

This sampling of AFBC development unit data is not intended to demonstrate a direct comparison with Westinghouse projections as much as to serve as an illustration of typical development unit performance near the base operating conditions of this study.

REFERENCES

C1. Keairns, D. L., et al., Fluidized-Bed Combustion Process Evaluation - Phase II - Pressurized Fluidized-Bed Coal Combustion Development, prepared for EPA by Westinghouse Research Laboratories, Pittsburgh, PA, 1975, EPA-650/2-75-027c, NTIS PB 246 116.

C2. Carls, E. L., et al., Reduction of Atmospheric Pollution by the Application of Fluidized Bed Combustion, Argonne National Laboratory, Argonne, IL, 1969, ANL ES-CEN-1001.

C3. Reduction of Atmospheric Pollution, National Coal Board, London, U.K., 1971, DHB 060971.

C4. Lange, H. B., and C. L. Chen., SO_2 Absorption in Fluidized Bed Combustion of Coal-Effect of Limestone Particle Size, The Babcock & Wilcox Company, Alliance, OH, 1978, EPRI FP-667.

C5. Fluidized Bed Combustion Development Facility and Commercial Utility AFBC Design Assessment, Quarterly prepared for EPRI, Babcock & Wilcox, Alliance, OH, April-June, 1978.

C6. Gordon, J. S., et al., Study of the Characterization and Control of Air Pollutants from a Fluidized-Bed Boiler - The SO_2 Acceptor Process, prepared for EPA, Pope, Evans and Robbins, Inc., Alexandria, VA, 1972, NTIS PB 229 242.

C7. Hammons, G. H., and A. Skopp, A Regenerative Limestone Process for Fluidized Bed Coal Combustion and Resulfurization, Final Report prepared under Contract CPA 70-19, Esso Research and Engineering Company, Linden, NJ, 1971.

C8. Jonke, A. A., et al., Reduction of Atmospheric Pollution by the Application of Fluidized-Bed Combustion, Argonne National Laboratory, Argonne, IL, 1970, ANL ES-CEN-1002.

C9. A Development Program on Pressurized Fluidized-Bed Combustion, Argonne National Laboratory, Argonne, IL, 1974, ANL ES-CEN-1007.

Table D1

COST FACTORS FOR ESTIMATING CAPITAL COSTS

Cost Base Year	Mid-1978
Site	Midwest
Direct Costs (DC)	Equipment cost + installation cost = installed cost
Indirect Costs (IC):	
Engineering services	10% of DC
Construction and field expenses	10% of DC
Contractor's fee	10% of DC
Start-up	2% of DC
Contingencies	20% of (DC + IC)
Turnkey Cost (TC)	DC + IC + contingencies
Working Capital (WC)	25% of total direct operating cost
Total Capital Investment	TC + WC

APPENDIX D

INDUSTRIAL-SIZED FLUIDIZED-BED BOILER COSTS

The capital and operating costs for the plant capacities under consideration have been estimated in mid-1978 dollars by assuming the cost factors given in Tables D1 and D2. Costs for the industrial FBC boiler modules and coal, sorbent, and spent solids handling systems were developed from Westinghouse cost files on related equipment and previous cost studies. Other sources of cost data are listed in Table D3. Costs for equipment items that also appear in conventional boiler plants were taken from the study by PedCo, used for conventional boiler cost comparisons, to provide greater consistency between conventional and FBC boiler cost comparisons.

The plant component capital and installation costs for the four plant sizes are summarized in Tables D4 and D5. The estimated costs for sorbent and spent solids handling systems are shown in Figure D1. Some of the estimated operating costs used in this study are shown in Table D6. These were again selected to be consistent with conventional boiler cost comparisons.

This cost information has been used to estimate the capital and operating costs summarized in Tables D7 to D18 for 500 μm particle sizes and in Tables D19 to D30 for 1000 μm particle sizes. In these tables the calcium-to-sulfur ratios are those reported in Section 6 of the main report, except that for the subbituminous coal no credit has been taken for sulfur capture by the coal-ash alkali. Thus, the calcium-to-sulfur ratios are significantly larger than in Section 6 for the subbituminous coal but with only a minor impact because of the low sulfur content of the subbituminous coal.

Table D2

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COST FACTORS FOR ESTIMATING ANNUAL COSTS[a]

Plant Capacity Factor, %	60
Direct Operating Cost (DOC):	
Direct labor, $ / man-hour	12.02[b]
Supervision, $ / man-hour	15.63[c]
Maintenance labor, $ / man-hour	14.63[b]
Electricity, mills/kWh	25.8[d]
Replacement parts (includes maintenance materials)	
Untreated water, $ / 1000 gal.	0.12
Process water, $ / 1000 gal.	0.15
Boiler feedwater, $ / 1000 gal.	1.00
Coal, high S, $\$/10^6$ Btu ($ / ton) (eastern)	0.74 (17.00)[e]
low S, $\$/10^6$ Btu ($ / ton) (eastern)	1.16 (29.00)[e]
low S, $\$/10^6$ Btu ($ / ton) (Wyoming)	0.42 (6.75)[e]
Limestone, $ / ton delivered (Western 90% CaL)	25.00
$ / ton delivered (Bussen quarry)	25.00
$ / ton delivered (Menlo quarry)	25.00
Sodium hydroxide, $ / ton (bulk, 50% basis, FOB works)	158.00[f]
Sodium carbonate, $ / ton (bulk, FOB works)	60.00[f]
Ammonia, $ / ton delivered	130.00[f]
Ammonium hydroxide, $ / ton (29.3% NH_3 basis, freight equalized)	173.00[f]
Solid waste disposal, $ / ton	8.00
Overhead Cost (OC):	
Payroll	30% of direct labor
Plant	26% of labor, parts and maintenance
Capital Charges (CC):	
G & A, taxes and insurance	4% of turnkey costs
Capital recovery	10.6% of turnkey costs
Interest on working capital where working capital is 25% of DOC	10% of working capital
Total Annual Costs (TAC)	DOC + OC + CC

[a] All costs are in June 1978 dollars.

[b] Engineering News-Record, June 29, 1978, pp. 52-53, average for Chicago, Cincinnati, Cleveland, Detroit, and St. Louis.

[c] Estimated at 30 percent over direct labor rate.

[d] EEI members publication for June 1978, average for Boston, Chicago, Indianapolis, Houston, San Francisco, and Los Angeles.

[e] Coal Outlook, July 18, 1978, issue, spot market prices.

[f] Chemical Marketing Reporter, June 19, 1978.

Table D3

COST SOURCES

Boiler	Westinghouse cost files
Primary Particulate Control Device	Neveril, R. B., J. U. Price and K. L. Engdahl, "Capital and Operating Costs of Selected Air Pollution Control Systems-II" *J. Air Pollution Control Association*, 28(9), 963 (1978)
Final Particulate Control Device	Neveril, R. B., J. U. Price and K. L. Engdahl, "Capital and Operating Costs of Selected Air Pollution Control Systems-I", *J. Air Pollution* Control Association, 28(8), 829 (1978)
Stack, deaerators, boiler feed pumps, condensate systems, water treating system, chemical feed, foundation and support, painting, electrical, building, etc.	PedCo, New Boiler Costs, report to EPA for preparation of Industrial Boiler ITAR, Section IV, October 1978.
Coal, sorbent and spent solids handling systems	Westinghouse cost files

Table D4

CAPITAL COSTS OF PLANTS EXCLUDING SORBENT AND SPENT SOLIDS HANDLING SYSTEMS*

Plant Thermal Input, 10^6 Btu/hr	30	75	150	200
Boiler (with fans, ducts, coal and sorbent feeders and instrumentation)	384,048	805,013	1,484,150	2,757,729
Primary Part. Control Device (cyclones)	19,500	31,250	62,500	86,400
Final Part. Control Device (baghouse)	86,350	190,525	361,575	473,225
Stack**	44,000	80,000	300,000	365,000
Deaerators**	5,200	15,900	21,000	29,000
Boiler Feed Pumps**	13,400	25,300	44,500	58,000
Condensate Systems**	7,700	8,700	9,200	16,300
Water Treating System**	8,000	15,000	18,000	20,000
Chemical Feed**	1,400	1,400	1,500	1,500
Coal Handling System	395,000	505,000	605,000	650,000
Spent Solids Withdrawal and Cooling System	28,700	57,800	98,170	122,300
TOTAL	993,298	1,735,888	3,005,595	4,579,454

86, 400

*Costs in mid-1978 dollars
**PedCo's estimate

Table D5

INSTALLATION COSTS OF PLANTS EXCLUDING SORBENT AND SPENT SOLIDS HANDLING SYSTEMS*

Plant Thermal Input, 10^6 Btu/hr	30	75	150	200
Boiler (including foundations, steel and insulation)**	134,417	281,755	519,453	965,205
Primary Part. Control Device**	6,825	10,940	21,875	30,240
Final Part. Control Device**	30,225	66,690	126,550	165,630
Stack***	10,000	20,000	50,000	60,000
Deaerators***	2,500	3,500	4,000	5,000
Boiler Feed Pumps***	3,000	5,500	7,000	8,000
Condensate System***	1,100	1,300	1,500	2,000
Water Treating System***	2,000	2,500	3,000	3,500
Chemical Feed***	800	1,500	1,500	1,500
Coal Handling System	Included with equipment cost			
Spent Solids Withdrawal and Cooling System**	10,045	20,230	34,360	42,810
Foundation and Support***	31,600	84,240	157,950	189,500
Piping***	41,000	58,500	81,900	93,600
Insulation	Included with boiler costs			
Painting***	5,900	8,200	11,700	11,700
Electrical***	30,000	75,000	150,000	160,000
Buildings***	140,400	234,000	409,500	444,600
TOTAL	449,812	873,855	1,580,288	2,183,285

*Mid 1978 dollars **Installation cost @ 35% of equipment cost ***PedCo's estimate

Table D6

OPERATING COSTS EXCLUDING COAL, SORBENT AND SOLIDS WASTE DISPOSAL COST AT 60% OPERATING LOAD*

Line words omitted

	Plant Size, MM Btu/hr			
	30	75	150	200
Direct Operating Cost				
Direct Labor	157,900	210,600	315,900	421,200
Supervision	68,500	136,900	136,900	136,900
Maintenance Labor	64,100	128,200	128,200	192,200
Replacement Parts	87,800	146,300	292,500	365,600
Chemicals	2,300	4,900	6,000	7,500
Process Water	4,700	9,500	18,800	25,200
Subtotal	385,300	636,400	898,300	1,148,600
Overhead (Payroll + Plant)	127,920	189,330	286,290	380,360

*Pedco estimate

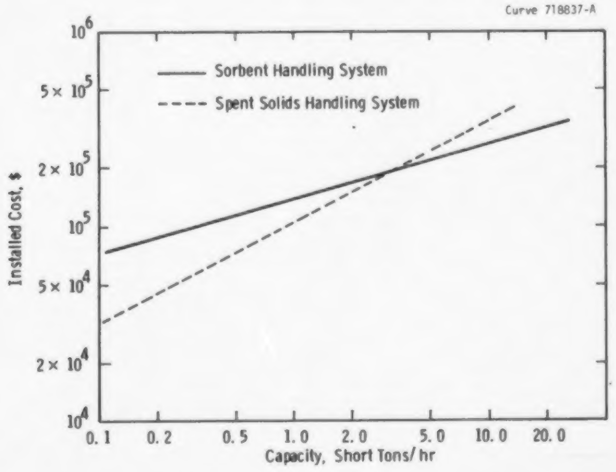

Figure D1 - Estimated Costs for Installed Sorbent and Spent Solids Handling Systems

Table D7

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	2516636.	2547854.	2567796.	2414662.	2428629.	2440606.	2414407.	2428070.	2439818.
WORKING CAPITAL	147890.	155737.	162433.	145552.	146915.	148374.	117986.	119293.	120691.
TOTAL CAPITAL COSTS	2664525.	2703591.	2730229.	2560214.	2575545.	2588980.	2532393.	2547363.	2560510.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	591558.	622948.	649733.	582208.	587662.	593497.	471946.	477173.	482765.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1101696.	1138429.	1168795.	1077223.	1084853.	1092583.	964168.	971521.	978968.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	103.	109.	115.	92.	93.	94.	93.	94.	95.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24509.	24365.	24001.	24903.	24880.	24805.	24273.	24251.	24179.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.55	8.89	9.27	8.23	8.30	8.38	7.56	7.62	7.70

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D8

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	2507669.	2534921.	2556437.	2414662.	2428629.	2440606.	2414407.	2428070.	2439818.
WORKING CAPITAL	145859.	152225.	158875.	145552.	146915.	148374.	117986.	119293.	120691.
TOTAL CAPITAL COSTS	2653528.	2687147.	2715312.	2560214.	2575545.	2588980.	2532393.	2547363.	2560510.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	583436.	608901.	635499.	582208.	587662.	593497.	471946.	477173.	482765.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1092062.	1122142.	1152546.	1077223.	1084853.	1092583.	964168.	971521.	978968.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	62.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	102.	107.	112.	92.	93.	94.	93.	94.	95.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24554.	24444.	24099.	24903.	24860.	24805.	24273.	24251.	24179.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.46	8.73	9.10	8.23	8.30	8.38	7.56	7.62	7.70

OSTS IN MID 1976 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D9

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	2495315.	2522077.	2544736.	2406258.	2420165.	2432718.	2406173.	2419787.	2432104.
WORKING CAPITAL	143276.	149032.	155461.	144857.	146055.	147417.	117320.	118468.	119773.
TOTAL CAPITAL COSTS	2638591.	2671109.	2700197.	2551115.	2566219.	2580135.	2523493.	2538255.	2551877.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	573103.	596127.	621845.	579427.	584218.	589666.	469281.	473873.	479094.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1079667.	1107173.	1136843.	1073146.	1080088.	1087505.	960234.	966928.	974078.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KWE	100.	104.	109.	91.	92.	93.	92.	93.	94.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24612.	24506.	24184.	24917.	24895.	24827.	24287.	24265.	24201.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.35	8.60	8.94	8.19	8.25	8.33	7.52	7.58	7.66

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

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Table D10

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - STRINGENT

OAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
OAL SULFUR CONTENT, WT%	3.50			.90			.60		
OAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
A/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
APITAL COSTS									
TOTAL TURNKEY	4472377.	4519201.	4548213.	4321911.	4341542.	4358044.	4322085.	4341233.	4357382.
WORKING CAPITAL	288011.	307630.	324371.	282167.	285576.	289223.	213254.	216521.	220016.
TOTAL CAPITAL COSTS	4760388.	4826831.	4872584.	4604078.	4627118.	4647267.	4535339.	4557754.	4577398.
PERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1152045.	1230521.	1297483.	1128669.	1142305.	1156893.	853015.	866082.	880063.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	2023143.	2110417.	2183289.	1977215.	1994058.	2011420.	1694694.	1710885.	1727572.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	258.	273.	287.	229.	232.	235.	232.	234.	237.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61273.	60912.	60003.	62258.	62199.	62011.	60683.	60627.	60447.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.28	6.59	6.92	6.04	6.10	6.17	5.31	5.37	5.44

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D11

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	4459188.	4499876.	4531112.	4321911.	4341542.	4358044.	4322085.	4341233.	4357382.
WORKING CAPITAL	282935.	298851.	315474.	282167.	285576.	289223.	213254.	216521.	220016.
TOTAL CAPITAL COSTS	4742123.	4798727.	4846586.	4604078.	4627118.	4647267.	4535339.	4557754.	4577398.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1131741.	1195403.	1261897.	1128669.	1142305.	1156893.	853015.	866082.	880063.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	2000406.	2071601.	2144317.	1977215.	1994058.	2011420.	1694694.	1710885.	1727572.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	254.	266.	280.	229.	232.	235.	232.	234.	237.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61385.	61110.	60248.	62258.	62199.	62011.	60683.	60627.	60447.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.20	6.45	6.77	6.04	6.10	6.17	5.31	5.37	5.44

OSTS IN MID 1976 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

CAUTION: Mark corrections in Sky Blue pencil ONLY.

Table D12

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	4441142.	4480750.	4513536.	4310313.	4329657.	4346873.	4310752.	4329637.	4346488.
WORKING CAPITAL	276477.	290867.	306941.	280429.	283424.	286829.	211588.	214458.	217721.
TOTAL CAPITAL COSTS	4717619.	4771617.	4820477.	4590742.	4613080.	4633702.	4522340.	4544094.	4564209.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1105907.	1163467.	1227763.	1121718.	1133695.	1147316.	846353.	857831.	870885.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	1971292.	2036073.	2106763.	1968396.	1983497.	1999972.	1686212.	1700734.	1716574.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KWE	249.	260.	273.	228.	230.	233.	231.	233.	235.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61529.	61265.	60461.	62293.	62237.	62068.	60717.	60663.	60501.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.10	6.32	6.63	6.01	6.06	6.13	5.28	5.33	5.40

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D13

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (150. MM Btu/hr THERMAL INPUT)

450 PGIG, 600 F STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	8365870.	8430691.	8469926.	8160136.	8185950.	8207284.	8160954.	8186071.	8206905.
WORKING CAPITAL	482398.	521636.	555117.	470710.	477527.	484822.	332882.	339416.	346406.
TOTAL CAPITAL COSTS	8848268.	8952326.	9025043.	8630846.	8663478.	8692105.	8493836.	8525487.	8553311.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1929590.	2086542.	2220467.	1882838.	1910110.	1939286.	1331529.	1357665.	1385625.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3485537.	3655877.	3798878.	3407579.	3439301.	3472322.	2842607.	2873063.	2904764.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	516.	546.	574.	458.	464.	470.	464.	469.	474.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111086.	110431.	108784.	112872.	112765.	112425.	110017.	109914.	109588.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.97	6.30	6.64	5.74	5.80	5.88	4.92	4.97	5.04

OSTS IN MID 1976 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

CAUTION: Mark corrections in blue pencil ONLY.

Table D14

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (150. MM Btu/hr THERMAL INPUT)

450 PGIG, 600 F STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	8347860.	8404014.	8446185.	8160136.	8185950.	8207284.	8160954.	8186071.	8206905.
WORKING CAPITAL	472246.	504077.	537324.	470710.	477527.	484822.	332882.	339416.	346406.
TOTAL CAPITAL COSTS	8820125.	8908091.	8983509.	8630846.	8663478.	8692105.	8493836.	8525487.	8553311.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1888982.	2016307.	2149295.	1882838.	1910110.	1939286.	1331529.	1357665.	1385625.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3441287.	3579991.	3722460.	3407579.	3439301.	3472322.	2842607.	2873063.	2904764.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	508.	533.	560.	458.	464.	470.	464.	469.	474.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111290.	110791.	109229.	112872.	112765.	112425.	110017.	109914.	109588.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.88	6.15	6.48	5.74	5.80	5.88	4.92	4.97	5.04

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D15

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (150. MM Btu/hr THERMAL INPUT)

450 PSIG, 600 F STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	8323397.	8377680.	8421828.	8145121.	8170335.	8192501.	8146315.	8170871.	8192523.
WORKING CAPITAL	459329.	488109.	520256.	467234.	473223.	480033.	329551.	335291.	341817.
TOTAL CAPITAL COSTS	8782725.	8865789.	8942084.	8612355.	8643557.	8672534.	8475867.	1506161.	8534340.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1837315.	1952435.	2081026.	1868935.	1892890.	1920132.	1318206.	1341163.	1367269.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3384753.	3510677.	3648928.	3391136.	3419371.	3450530.	2826813.	2853929.	2883849.
PERFORMANCE									
BOILER EFFICIENCY, %	64.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KWE	498.	520.	547.	456.	460.	466.	461.	466.	471.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111550.	111072.	109614.	112935.	112834.	112528.	110077.	109980.	109687.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.77	6.01	6.33	5.71	5.77	5.83	4.89	4.94	5.00

OSTS IN MID 1978 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D16

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	11141702.	11216233.	11260915.	10906340.	10935389.	10959217.	10907547.	10935782.	10959031.
WORKING CAPITAL	630913.	683231.	727872.	615329.	624420.	634145.	431560.	440272.	449592.
TOTAL CAPITAL COSTS	11772615.	11899463.	11988787.	11521669.	11559809.	11593362.	11339106.	11376053.	11408622.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2523654.	2732923.	2911489.	2461317.	2497680.	2536582.	1726239.	1761086.	1798367.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4593794.	4819176.	5008730.	4495536.	4537048.	4580402.	3742257.	3782098.	3823705.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	688.	729.	766.	611.	618.	626.	618.	625.	633.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139562.	138739.	136669.	141805.	141671.	141244.	138218.	138089.	137680.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.26	6.61	6.97	6.03	6.09	6.17	5.15	5.21	5.28

OSTS IN MID 1976 DOLLARS
ORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
ORBENT PARTICLE SIZE - 500. MICRONS

Table D17

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

OAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
OAL SULFUR CONTENT, WT%	3.50			.90			.60		
OAL HHV, BTU/LB	11800.			13800.			9600.		
REMOVAL OF SO2 RELEASED	65.00			84.67			84.00		
O2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
ORBENT TYPE	I	II	III	I	II	III	I	II	III
A/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
APITAL COSTS									
TOTAL TURNKEY	11121141.	11185595.	11233586.	10906340.	10935389.	10959217.	10907547.	10935782.	10959031.
WORKING CAPITAL	617377.	659819.	704148.	615329.	624420.	634145.	431560.	440272.	449592.
TOTAL CAPITAL COSTS	11738518.	11845414.	11937734.	11521669.	11559809.	11593362.	11339106.	11376053.	11408622.
PERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2469510.	2639276.	2816593.	2461317.	2497680.	2536582.	1726239.	1761086.	1798367.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4535294.	4718715.	4907472.	4495536.	4537048.	4580402.	3742257.	3782098.	3823705.
ERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KWE	678.	710.	747.	611.	618.	626.	618.	625.	633.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139817.	139191.	137228.	141805.	141671.	141244.	138218.	138089.	137680.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.17	6.45	6.80	6.03	6.09	6.17	5.15	5.21	5.28

STS IN MID 1978 DOLLARS
RBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
RBENT PARTICLE SIZE - 500. MICRONS

Table D18

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - MODERATE

OAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
OAL SULFUR CONTENT, WT%	3.50			.90			.60		
OAL HHV, BTU/LB	11800.			13800.			9600.		
REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
O2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
ORBENT TYPE	I	II	III	I	II	III	I	II	III
A/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
APITAL COSTS									
TOTAL TURNKEY	11093222.	11155382.	11205569.	10889557.	10917823.	10942536.	10891203.	10918701.	10943218.
WORKING CAPITAL	600155.	638528.	681392.	610695.	618680.	627761.	427119.	434771.	443073.
TOTAL CAPITAL COSTS	11693377.	11793910.	11886960.	11500252.	11536503.	11570297.	11318321.	11353472.	11386291.
PERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2400620.	2554113.	2725567.	2442780.	2474720.	2511043.	1708474.	1739084.	1773893.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4460605.	4627011.	4810080.	4474085.	4510950.	4551789.	3721662.	3757051.	3796251.
ERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KWE	664.	694.	729.	607.	614.	621.	615.	621.	628.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	140145.	139544.	137712.	141885.	141757.	141373.	138294.	138172.	137804.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.06	6.31	6.65	6.00	6.05	6.13	5.12	5.17	5.24

STS IN MID 1978 DOLLARS
RBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
RBENT PARTICLE SIZE - 500. MICRONS

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Table D19

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	3.45	4.32	8.46	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	2531306.	2569494.	2618644.	2420593.	2438466.	2463903.	2420223.	2437702.	2462615.
WORKING CAPITAL	151227.	161770.	180623.	146032.	147947.	151637.	118446.	120282.	123818.
TOTAL CAPITAL COSTS	2682533.	2731263.	2799267.	2566624.	2586413.	2615540.	2538669.	2557984.	2586433.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	604910.	647079.	722492.	584126.	591789.	606548.	473785.	481128.	495272.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1117523.	1166322.	1250796.	1080056.	1090520.	1109362.	966902.	976981.	995116.
PERFORMANCE									
BOILER EFFICIENCY, %	83.47	82.74	79.84	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	94.	96.	100.	90.	90.	91.	91.	91.	92.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24370.	24157.	23309.	24883.	24844.	24680.	24254.	24216.	24060.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.72	9.19	10.21	8.26	8.35	8.55	7.58	7.68	7.87

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

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Table D20

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.94	3.67	7.38	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	2518991.	2553912.	2602316.	2420593.	2438466.	2463903.	2420223.	2437702.	2462615.
WORKING CAPITAL	148186.	157079.	174205.	146032.	147947.	151637.	118446.	120282.	123818.
TOTAL CAPITAL COSTS	2667177.	2710991.	2776521.	2566624.	2586413.	2615540.	2538669.	2557984.	2586433.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	592744.	628315.	696821.	584126.	591789.	606548.	473785.	481128.	495272.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1103255.	1144814.	1222100.	1080056.	1090520.	1109362.	966902.	976981.	995116.
PERFORMANCE									
BOILER EFFICIENCY, %	83.76	83.15	80.54	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	93.	95.	98.	90.	90.	91.	91.	91.	92.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24455.	24277.	23515.	24883.	24844.	24680.	24254.	24216.	24060.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.58	8.97	9.89	8.26	8.35	8.55	7.58	7.68	7.87

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D21

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (30. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.45	3.13	6.41	2.25	2.90	5.98	2.25	2.90	5.98
CAPITAL COSTS									
TOTAL TURNKEY	2505611.	2539475.	2586385.	2411789.	2429087.	2453559.	2411600.	2428527.	2452503.
WORKING CAPITAL	145186.	153114.	168412.	145245.	146883.	150047.	117693.	119262.	122294.
TOTAL CAPITAL COSTS	2650796.	2692590.	2754797.	2557034.	2575970.	2603606.	2529293.	2547789.	2574797.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	580743.	612458.	673648.	580981.	587532.	600187.	470770.	477048.	489176.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1089000.	1126452.	1196021.	1075547.	1084787.	1101331.	962553.	971459.	987391.
PERFORMANCE									
BOILER EFFICIENCY, %	84.02	83.45	81.14	85.29	85.17	84.69	83.13	83.01	82.56
AUXILIARY POWER, KWE	93.	94.	97.	90.	90.	91.	91.	91.	92.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24530.	24365.	23690.	24901.	24866.	24727.	24271.	24238.	24104.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.45	8.80	9.61	8.22	8.30	8.47	7.55	7.63	7.79

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D22

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	3.45	4.32	8.46	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	4493792.	4551535.	4625067.	4330056.	4355328.	4391126.	4330053.	4354695.	4389667.
WORKING CAPITAL	296356.	322712.	369845.	283366.	288156.	297380.	214403.	218993.	227833.
TOTAL CAPITAL COSTS	4790148.	4874247.	4994912.	4613423.	4643484.	4688506.	4544456.	4573688.	4617500.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1185424.	1290848.	1479379.	1133465.	1152622.	1189520.	857611.	875970.	911330.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	2060484.	2176973.	2380953.	1983320.	2006646.	2049692.	1700569.	1722985.	1764335.
PERFORMANCE									
BOILER EFFICIENCY, %	83.47	82.74	79.84	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	234.	239.	249.	225.	226.	227.	227.	228.	230.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	60926.	60394.	58274.	62207.	62110.	61701.	60634.	60541.	60149.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.43	6.86	7.77	6.07	6.15	6.32	5.34	5.41	5.58

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D23

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.94	3.67	7.38	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	4475616.	4528069.	4600106.	4330056.	4355328.	4391126.	4330053.	4354695.	4389667.
WORKING CAPITAL	288753.	310984.	353801.	283366.	288156.	297380.	214403.	218993.	227833.
TOTAL CAPITAL COSTS	4764369.	4839053.	4953907.	4613423.	4643484.	4688506.	4544456.	4573688.	4617500.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1155010.	1243937.	1415203.	1133465.	1152622.	1189520.	857611.	875970.	911330.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	2026656.	2125463.	2311528.	1983320.	2006646.	2049692.	1700569.	1722985.	1764335.
PERFORMANCE									
BOILER EFFICIENCY, %	83.76	83.15	80.54	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	233.	237.	246.	225.	226.	227.	227.	228.	230.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61137.	60691.	58788.	62207.	62110.	61701.	60634.	60541.	60149.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.31	6.66	7.48	6.07	6.15	6.32	5.34	5.41	5.58

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D24

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(75. MM Btu/hr THERMAL INPUT)

150 PSIG SATURATED STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.45	3.13	6.41	2.25	2.90	5.98	2.25	2.90	5.98
CAPITAL COSTS									
TOTAL TURNKEY	4455964.	4506409.	4575838.	4317847.	4342060.	4376255.	4318127.	4341755.	4375170.
WORKING CAPITAL	281252.	301073.	339317.	281401.	285495.	293404.	212519.	216443.	224023.
TOTAL CAPITAL COSTS	4737215.	4807482.	4915155.	4599247.	4627555.	4669660.	4530646.	4558197.	4599192.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1125006.	1204294.	1357269.	1125602.	1141979.	1173617.	850076.	865770.	896090.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	1993032.	2081667.	2248603.	1973478.	1993799.	2031221.	1691104.	1710641.	1746597.
PERFORMANCE									
BOILER EFFICIENCY, %	84.02	83.45	81.14	85.29	85.17	84.69	83.13	83.01	82.56
AUXILIARY POWER, KWE	231.	235.	243.	224.	225.	227.	227.	228.	229.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61325.	60912.	59226.	62251.	62165.	61817.	60677.	60594.	60261.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.18	6.50	7.22	6.03	6.10	6.25	5.30	5.37	5.51

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D25

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (150. MM Btu/hr THERMAL INPUT)

450 PGIG, 600 F STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	3.45	4.32	8.46	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	8394908.	8475327.	8576928.	8170637.	8204035.	8251159.	8171203.	8203690.	8249629.
WORKING CAPITAL	499087.	551799.	646065.	473108.	482686.	501135.	335181.	344360.	362040.
TOTAL CAPITAL COSTS	8893995.	9027126.	9222992.	8643745.	8686721.	8752294.	8506383.	8548050.	8611669.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1996349.	2207196.	2584258.	1892431.	1930745.	2004540.	1340723.	1377440.	1448161.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3558204.	3786064.	4187386.	3418945.	3463093.	3545613.	2853526.	2895905.	2975101.
PERFORMANCE									
BOILER EFFICIENCY, %	83.47	82.74	79.84	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	469.	479.	498.	449.	451.	455.	455.	457.	460.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	110457.	109492.	105648.	112779.	112603.	111861.	109928.	109759.	109048.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.13	6.58	7.54	5.77	5.85	6.03	4.94	5.02	5.19

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D26

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER
(150. MM Btu/hr THERMAL INPUT)

450 PSIG, 600 F STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	65.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.94	3.67	7.38	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	8370051.	8442743.	8541886.	8170637.	8204035.	8251159.	8171203.	8203690.	8249629.
WORKING CAPITAL	483880.	528343.	613977.	473108.	482686.	501135.	335181.	344360.	362040.
TOTAL CAPITAL COSTS	8853931.	8971086.	9155863.	8643745.	8686721.	8752294.	8506383.	8548050.	8611669.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1935521.	2113374.	2455906.	1892431.	1930745.	2004540.	1340723.	1377440.	1448161.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3492226.	3685139.	4050709.	3418945.	3463093.	3545613.	2853526.	2895905.	2975101.
PERFORMANCE									
BOILER EFFICIENCY, %	83.76	83.15	80.54	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	466.	474.	492.	449.	451.	455.	455.	457.	460.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	110840.	110031.	106581.	112779.	112603.	111861.	109928.	109759.	109048.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.99	6.37	7.23	5.77	5.85	6.03	4.94	5.02	5.19

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D27

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (150. MM Btu/hr THERMAL INPUT)

450 PSIG, 600 F STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.45	3.13	6.41	2.25	2.90	5.98	2.25	2.90	5.98
CAPITAL COSTS									
TOTAL TURNKEY	8343275.	8412749.	8507903.	8154763.	8186495.	8231237.	8155734.	8186623.	8230250.
WORKING CAPITAL	468878.	508522.	585010.	469176.	477365.	493184.	331413.	339260.	354420.
TOTAL CAPITAL COSTS	8812153.	8921271.	9092912.	8623939.	8663859.	8724421.	8487147.	8525883.	8584670.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1875513.	2034088.	2340038.	1876705.	1909458.	1972734.	1325652.	1357040.	1417680.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3426809.	3599492.	3926983.	3400508.	3438713.	3510103.	2835820.	2872503.	2941029.
PERFORMANCE									
BOILER EFFICIENCY, %	84.02	83.45	81.14	85.29	85.17	84.69	83.13	83.01	82.56
AUXILIARY POWER, KWE	463.	470.	486.	448.	450.	453.	454.	456.	459.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111180.	110432.	107374.	112860.	112704.	112073.	110005.	109855.	109251.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.86	6.20	6.96	5.73	5.81	5.96	4.90	4.97	5.12

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D28

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - STRINGENT

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	3.45	4.32	8.46	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	11174807.	11267497.	11384227.	10918054.	10955718.	11008773.	10918969.	10955569.	11007241.
WORKING CAPITAL	653166.	723449.	849136.	618527.	631298.	655897.	434624.	446863.	470437.
TOTAL CAPITAL COSTS	11827974.	11990945.	12233363.	11536581.	11587017.	11664670.	11353593.	11402432.	11477678.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2612665.	2893795.	3396545.	2474108.	2525194.	2623587.	1738497.	1787454.	1881748.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4689864.	4991554.	5523915.	4510357.	4568218.	4676818.	3756489.	3812013.	3916209.
PERFORMANCE									
BOILER EFFICIENCY, %	83.47	82.74	79.84	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	625.	639.	665.	599.	601.	606.	607.	609.	614.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	138771.	137558.	132730.	141688.	141468.	140535.	138106.	137894.	137001.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.43	6.90	7.92	6.06	6.14	6.33	5.18	5.26	5.44

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D29

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - INTERMEDIATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	65.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.94	3.67	7.38	2.84	3.56	7.19	2.84	3.56	7.19
CAPITAL COSTS									
TOTAL TURNKEY	11146368.	11229985.	11343710.	10918054.	10955718.	11008773.	10918969.	10955569.	11007241.
WORKING CAPITAL	632890.	692175.	806352.	618527.	631298.	655897.	434624.	446863.	470437.
TOTAL CAPITAL COSTS	11779259.	11922160.	12150062.	11536581.	11587017.	11664670.	11353593.	11402432.	11477678.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2531561.	2768698.	3225408.	2474108.	2525194.	2623587.	1738497.	1787454.	1881748.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4602580.	4857854.	5342585.	4510357.	4568218.	4676818.	3756489.	3812013.	3916209.
PERFORMANCE									
BOILER EFFICIENCY, %	83.76	83.15	80.54	85.22	85.09	84.53	83.07	82.94	82.40
AUXILIARY POWER, KWE	621.	632.	656.	599.	601.	606.	607.	609.	614.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139252.	138236.	133901.	141688.	141468.	140535.	138106.	137894.	137001.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.29	6.69	7.59	6.06	6.14	6.33	5.18	5.26	5.44

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS

Table D30

ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFB INDUSTRIAL BOILER (200. MM Btu/hr THERMAL INPUT)

750 PSIG, 750 F STEAM

SO2 CONTROL LEVEL - MODERATE

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.45	3.13	6.41	2.25	2.90	5.98	2.25	2.90	5.98
CAPITAL COSTS									
TOTAL TURNKEY	11115781.	11195495.	11304457.	10900279.	10935934.	10986176.	10901666.	10936337.	10985280.
WORKING CAPITAL	612888.	665746.	767729.	613285.	624203.	645295.	429601.	440063.	460277.
TOTAL CAPITAL COSTS	11728668.	11861241.	12072186.	11513564.	11560137.	11631471.	11331266.	11376401.	11445556.
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2451550.	2662984.	3070918.	2453140.	2496811.	2581179.	1718403.	1760254.	1841107.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4516103.	4744461.	5178502.	4486269.	4536237.	4630050.	3733366.	3781325.	3871345.
PERFORMANCE									
BOILER EFFICIENCY, %	84.02	83.45	81.14	85.29	85.17	84.69	83.13	83.01	82.56
AUXILIARY POWER, KWE	617.	627.	648.	598.	600.	604.	606.	608.	612.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139680.	138739.	134898.	141790.	141594.	140801.	138203.	138015.	137256.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.15	6.51	7.30	6.02	6.10	6.26	5.14	5.21	5.37

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 1000. MICRONS